National
Scan

May 31, 1950

Dr. H. F. Johnstone
Head, Division of Chemical Engineering
Department of Chemistry
University of Illinois
Urbana, Illinois

Dear Sir:

The attached report of a pre-pilot plant design for preparation of gasoline by the Fischer-Tropsch process has been prepared by the graduate plant design class.

For greater efficiency, the class was subdivided into three groups, responsible respectively for feed preparation, reactor design and product separation. Later, a fourth group responsible for the plant layout and the preparation of the final report was formed. Members of the groups and group chairmen are as follows:

Feed Preparation:	Bollinger, Coldren, Kunstman, Robb, Tung. Chairman: Robb.
Reactor Design:	Feild, Hofto, McKinley, Timmerhaus, White, Chairman: Feild.
Product Separation:	Caskey, Chiang, Houvouras, Junk, Koeller, Schueler. Chairman: Schueler.
Plant Layout:	Bollinger, Caskey, Houvouras, McKinley, Tung, White. Chairman: White.

It is estimated that gasoline can be produced at a cost of 14.7 cents per gallon. 219,888 lbs per day of acids, aldehydes, ketones and oil-soluble chemicals are produced. Their sale is expected to defray their separation cost and refining cost. Total plant investment is $45,587,125. Annual operating costs are estimated as $10,130,381.

The process is non-competitive with gasoline from crude at present crude prices.

Very truly yours,

Edward A. White

Edward A. White
Chairman: Plant layout
and final report group.

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PRE-PILOT PLANT DESIGN

of

A FISCHER-TROPSCH GASOLINE PLANT

CHEMICAL ENGINEERING 468

Spring, 1950

Sarangapani, Mar 31, 1961

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CHEMICAL ENGINEERING 488

Spring, 1950

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Synthesis Gas Preparations II

Reactor III

Product Separation IV

Flow Sheets and Layouts (envelope on back cover)

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Feed Promotion - TT

General - I

GENERAL SECTION

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SUMMARY

At the request of Dr. H. F. Johnstone, Head, Division of Chemical Engineering, a Fischer-Tropsch plant to produce 5,000 barrels of gasoline from natural gas has been designed. In addition to the gasoline, 219,888pounds per day of oxygenated chemicals are produced as side reaction products.

The cost of production of gasoline is 14.7 cents per gallon. Total plant investment is $45,587,125. Annual operating costs, including cost of natural gas and other materials, are computed as $10,130,381. Depreciation is based on 10 year amortization with the exception of equipment in very severe service amortized over shorter periods. The value of the oxygenated products is expected to defray the cost of separation.

The process may be summarized as follows. Natural gas is compressed to 300 psia and treated to remove hydrogen sulfide and then partially oxidized with tonnage oxygen to carbon monoxide and hydrogen. The carbon monoxide-hydrogen mixture (synthesis gas) is reacted in iron catalyst fluidized-bed reactors to yield hydrocarbons and oxygenated products. The reactor operates at 300psia and 600° F.

The product stream form the reactor is sent to the product separation unit where it is condensed. Hydrocarbons,

oil and water soluble chemicals are removed and the unreacted synthesis gas recycled. Sufficient gas is bled to keep the inerts concentration in the reactor feed stream constant. Liquid hydrocarbons are sent to the refinery for processing; gaseous hydrocarbons are polymerized to gasoline in the polymerization plant.

Plant location has been tentatively selected as Baton Rouge, Louisiana. The Mississippi river provides an inexhaustible source of cooling water and cheap transportation of the products of the plant. Proved natural gas reserves in the state are 27×10^{12} cubic feet, the second largest state reserve in the United States. The current price for this natural gas is 3.9 cent per thousand cubic feet at the well. Other factors that favor this location are: mild climate allowing cheaper open building construction; a plentiful, economic labor supply; a large concentration of chemical industry in the region for an outlet for the chemicals production and an urban area simplifying the problem of attracting and holding key and skilled personnel.

The high production cost makes construction of this plant unjustifiable on an economic basis. Construction of the plant might possibly be justified on a strategic basis, though it seems unlikely that 5,000 barrels of gasoline per day would justify such an expenditure of material and man-hours.

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GENERAL CONSIDERATIONS

The plant location is determined by several factors. The primary consideration is that the plant be located at or near a source of plentiful, cheap natural gas. Texas and Louisiana have the largest proved gas resources in the country. Texas has proved reserves of 99 trillion cubic feet while Louisiana has proved reserves of 27 trillion cubic feet. The well price in Texas averages 4.5 cents per 1,000 cubic feet while the price in Louisiana averages 3.9 cents per 1000 cubic feet. The state with the next largest reserve is Kansas with 14 trillion cubic feet at an average price at the well of 5.0 cents per M cubic feet.

The cooling water requirement is 5,000, 000 gallons per hour. It is highly improbable that conservation regulations would allow even the makeup for such a requirement to be drawn from the ground water table. Furthermore, cooling towers to carry such a load would be very costly. The obvious solution to this problem is to locate the plant on a river from which cooling water might be withdrawn and returned.

Location of the plant on a river has the additional advantage of allowing products to be shipped by water, assuming the river one selects is navigable. Barge and tanker provide the cheapest transportation. Adequate rail service should be available at the site, also.

The consideration of markets is necessary. Assuming that water transportation is available, it is posssible to ship cheaply to the Gulf South market, the Midwest market and to the East Coast. However, the volume of chemicals production is such that it would probably prove impractical to ship these in barge or tanker quantities. Therefore, location near an area with chemical development is prudent since the chemicals will be used mainly in chemical manufacturing and processing.

Labor is a major consideration. It is doubtful that at this time any given area is freer from the strike threat than any other. Unions are now nationally organized and appear to be uniformly militant as regards geography. Wage scales, labor supply and housing must then be considered. If feasible, the plant should be located near a town large enough so that the plant will not affect the housing situation measurably and large enough to provide adequate schools, churches, shopping facilities and recreation facilities. Obviously, a lower prevailing wage scale is advantageous.

Having stated the principal requirements, it is now indicated that the proposed locations be examined in the light of these. On the basis of natural gas supply, Louisiana is the obvious choice since it is the cheapest source of natural gas. The reserve is adequate for this plant certainly. However, state conservation regulations, if any, and the political aspects of conservation in the state should be carefully checked. The possibility of having to

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abandon a $50 million investment due to local demagoguery might exist.

Having selected Louisiana, at least tentatively, location along the Mississippi is indicated. The water supply is certainly inexhaustible and the river is navigable. One might locate in the New Orleans area or in the Baton Rouge area. However, the New Orleans area seems to offer little advantage if any over the Baton Rouge area. Furthermore, land around the Baton Rouge area is cheaper and much more solid. Baton Rouge is large enough to produce adequate labor and to provide housing, schools, etc. There is considerable chemical industry in the area and it is close to the Texas Gulf chemical area.

Having selected the plant location, the plant layout should be considered. In the Baton Rouge area, very little grading and filling will be necessary to prepare the plant site. Rail spurs have been placed so that all units will be available by rail. All but the largest units of the process equipment will be shipped in by rail. Having the site of these units close to rail spurs will reduce construction difficulties. The plant is located two miles from the river to reduce flood danger in case the levee fails. Future expansion area is located between the river and the plant. Due to the integrated nature of the plant

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as designed, expansion will be effected by duplication of all the present units with the possible exception of the power plant.

Shops, laboratories and administration buildings are located at a distance from the plant area. In the event of fire or explosion, personnel in these buildings will not be endangered. For the same reason, the fire department is located out of the plant area, so that in case of fire it will not be immediately nullified. First aid stations are located close to the various units. The most probable serious injuries in the operation will be burns. In this case it is advisable to bring first aid to the man, rather than the man to first aid. For minor injuries, men will be able to report to first aid and get back to their jobs with a minimum of time lost.

Roads are provided past the various units for maintenance truck and service trucks. The roads will be gravel initially and may then be paved later with asphalt from the refinery unit.

Laboratories will provide facilities for control analysis for all units, as well as pilot plant and research laboratories. Research and pilot plant work will, at least initially, be concerned with better separation methods and improvement of reactor yields.

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[illegible]

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[illegible]

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The plant presents a major safety hazard. This is the danger of fire and/or explosion from combustion of the synthesis gas preparation and reactor streams. Some of the separation plant streams will be quite flammable. A second, lesser hazard exists as carbon monoxide poisoning.

Fire and/or explosion will arise from leaks in the synthesis gas and reactor sections. Units that should be given very frequent periodic inspection are: tubes in the natural gas preheater in the feed preparation unit and reactor feed lines in the reactor section. In the preheater, the tubes are in very severe service-- 300 psi at 1200°F. The consequences of failure of one of these tubes would at least be damage to the furnace. The reactor feed line is under very severe service conditions also. Here the danger is catalyst erosion. Obviously, other units are hazardous, but these present the greatest hazard.

Instrumentation for these hazards is as follows: a temperature recorder in the furnace sounds an alarm and begins shutdown of the units when the temperature rises above 2600°F in the furnace (normal operating temperature in the furnace is 2400°F.). In the reactor line, small holes are drilled from the outside to the minimum allowable wall thickness in the line. Pressure taps installed on these holes lead to a pressure signal. When a pressure is transmitted to the pressure recorder, indicating that the wall thickness is the minimum safe thickness an alarm is sounded. The flow to the reactor is then shut off and the unit replaced.

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Other hazards will be dealt with according to established practice. These include clearances, ventilation water supplies, and process hazards.

Overall Cost Estimation -

A. Total Investment The total overall investment estimation is based on the total basic equipment cost Some of the designers of the equipment have provided detailed cost data. But since this is only a preliminary cost estimate great accuracy cannot be expected and their detailed cost data are not used in the overall estimation The estimation is base on the method given on page 426 of Vilbrandt's "Chemical Engineering Plant Design"

Basic Equipment Cost		
Synthesis Gas Plant	$	3,492,253
Fischer Tropsch reactor cost		965,160
Separation Plant		2,726,340
Total	$	7,178,653
Process piping ($\frac{21}{34}$ × Basic equipment Cost)		4,433,877
Electrical installations ($\frac{4}{34}$ × " " ")		844,548
Buildings and building service ($\frac{5}{34}$ " " " ")		1,055,685
Yard improvements ($\frac{3}{34}$ × " " ")		633,411
Home, field office expenses, contingency, interest insurance, and taxes ($\frac{23}{34}$ × " " ")		4,856,151
Total		19,002,325
Filtration plant		5,250,000
Oxygen Plant		10,000,000
Steam Plant		11,300,000
Total equipment investment	$	45,552,325
Fisher Tropsch catalyst initial cost		10,000
Water Shift catalyst initial cost		4,800
Diethanolamine initial cost		20,000
Total Investment	$	45,587,125

B. Annual Cost

Fixed cost

Depriciation of equipment (Based on 10 yrs, amortization)	[illegible]
Operational personnel (400 employees at average annual payment of $ 4,000 per employee)	1,600,000
Interest of Capital (at 6%)	[illegible]

Operational Cost

Diethanolamine replacement	50,038
Natural gas consumption (at 3.9 cents per 1000 standard cu ft)	[illegible]
Fischer Tropsch catalyst replacement	1,200
Water shift reaction catalyst replacement	1,600
	$ 10,130,381

C Cost per gallon of crude product

Number of barrels produced per year $= 5000 \times 365 \times \frac{90}{100} = 1,642,500$

Cost per gallon $= \frac{10,130,381}{1,642,500 \times 42} = \$\,0.147$

Total Steam Requirement

	650 p.s.i.g. steam	50 p.s.i.g. steam
Oxygen plant	440,000 lb/hr	
Filtration plant	36,500 "	
Synthesis Gas Plant	189,870 "	51,855 lb/hr
Fischer Tropsch Reactor unit	48,220 "	
Separation Plant	7,000 "	180,000 "
	721,590 "	231,855

Total Steam consumption 953,445 lb/hr

Total Natural Gas Consumption

Feed	61,920,000	Standard cu ft/day
Fuel for preheater	3,897,600	"
Fuel for steam plant	27,000,000	"
Total	92,817,600	"

Cooling Water

Cooling water required for the entire plant is 5,100,000 gallons per hour. This includes heat exchange for all sections, process and turbine condensers and boiler makeup water. A breakdown of individual section requirements is given below.

	gallons per hour
Oxygen compressor	22,600
Natural gas compressor	0
Natural gas scrubber	93,500
Hydrogen sulfide removal (cooler & condenser)	12,000
Synthesis Gas Scrubber #1	72,000
Synthesis Gas Scrubber #2	16,300
Carbon Dioxide Removal (cooler & condenser)	41,500
Oxygen Plant	1,250,000
Separations: intercoolers	494,000
scrubbers	160,000
absorbers	320,000
Turbine condensers	2,610,000
Boiler make-up (zeolite treated)	7,500
	5,100,200

Steam Plant

A steam plant for the production of steam for the Fischer-Tropsch plant, refinery and polymerization units will be purchased. The capacity of this installation is to be 1,000,000 pounds of steam per hour. This plant will have an electrical generation capacity of 25,000 kilowatts. Excess capacity is to be disposed of as electrical power. Operating data and costs given below are based on data on steam boiler and power plants (Chemical Engineers Hanbook, J. H. Perry, Ed., 3rd Ed., 1633, McGraw-Hill Book Co. Inc., New York, 1950).

Capacity, lb/hr	1,000,000
Heat added to steam, BTU/lb	1,100
Total heat added, BTU/hr	1.1×10^9
Boiler efficiency, %	85
Natural gas consumption, Cu.ft/day	27,000,000
Investment	$11,300,000
Fixed Annual Costs	*** 1,355,000
Labor, supervision	455,000
Maintenance, m&l	130,800
Supplies	23,800
Annual fuel cost	346,000
Total Annual Operating Cost	$1,964,600

[illegible]

[illegible] the water [illegible]

[illegible]

Following [illegible] is given on the [illegible] with [illegible] "[illegible]"

The [illegible] cost of the [illegible] is based on the costs of similar [illegible]. The requirement is 5,000 [illegible]

[illegible]

Flow Diagram for Rapid Sand Filtration Plant

Feed Preparation - II

SYNTHESIS GAS PREPARATION

Division of Chemical Engineering
University of Illinois
Urbana, Illinois

Plant Design Group

SYNTHESIS GAS PREPARATION

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Synthesis Gas Preparation

Introduction — 352 Notes

There are three possible industrial processes for the generation of synthesis feed gas from natural gas: (1) Use of a reducible metal oxide as an oxygen function medium for the partial oxidation of natural gas (2) Reforming of natural gas with steam and carbon dioxide and (3) Incomplete combustion of natural gas with oxygen. Since large scale process oxygen is commercially available now, the cost of the third process becomes comparable with the second and lower than the first. The incomplete combustion natural gas is chosen for the present plant because of the cost factor and the relatively simple high pressure equipment involved.

In this process the natural gas from the well and the 70% process oxygen are compressed to [illegible] psi. The sulphur content of the natural gas of the present location is reduced from [illegible] to [illegible] than [illegible] by the Sulfotol process. The slip is [illegible] since sulphur is very harmful to the catalyst of the Fischer Tropsch reactor and is undesirable in the finished gasoline.

Both the natural gas and oxygen are [illegible] in a [illegible] the natural gas to [illegible] and the oxygen to 590°F. The natural gas and oxygen are fed into the synthesis gas generator at a rate [illegible] to [illegible] at a temperature of [illegible] F. The temperature [illegible] at the reaction zone is about 2400°F. The exactly [illegible] reaction produces H_2 and CO in the ratio of 1.7H [illegible] with a [illegible] amount of [illegible] SH_2 and O_2. The gas is cooled to about 600°F in a waste heat recovery system in the upper section of the reaction vessel. This heat is used to generate steam at 680 psi for plant use. The gas then pass through a carbon scrubber to remove any water found and to cool them to 210°F.

Around 3% of the cooled gas is sent to a shift

reactor which operates at 590°F and uses ferric oxide as a catalyst. The gas is injected into the reactor along with some process steam. The products of the reactor, CO_2 and H_2, are cooled in a venturi scrubber. A CO_2 absorption unit reduces the CO_2 content to practically zero. Part of this H_2 rich stream is fed into the main synthesis gas stream to raise the H_2 - CO ratio to 1.9 as required by the Fischer Tropsch reactor. The other part of the H_2 rich stream is used for Fischer Tropsch catalyst regeneration.

Oxygen Plant

A preliminary survey of 95 % oxygen producing facilities disclosed that several 1000 ton/ day plants have been designed and two such units have been built. Patents and designs for oxygen producing equipment are in the hands of a few companies who wish to sell or lease complete plants. Therefore it has been decided that facilities for producing 2000 tons/day of 95 % oxygen will be purchased from a suitable company.

On the basis of a careful survey of the published data, the following cost and operating data are recommended for use in this report:

Total Investment (cost of plant, completely installed)	$ 10,000,000
Steam requirement (including compressors and electric power)	440,000 #/hr.
Water requirement	1,250,000 gal/hr.

This plant will be built in two 1000 ton/day units. The oxygen will be supplied to the synthesis gas section in 95 % purity at 1 atm. pressure and a temperature of 50 °F. The cost of the oxygen , based on a complete cost calculation including all fixed charges and operating costs for the oxygen plant, is $4.00 per ton.

References

Anonymous Chem. Eng. 54 ,No. 1, 123-30 (1947)

Anonymous Chem. Eng. 54, No. 3, 126-41 (1947)

Downs, C. R. Chem. Eng. 55, No. 8, 113-7,121 (1948)

Newman, L. L. Ind. Eng. Chem. 40, 559-82 (1948)

Coldren, C. L. Chem. Eng. Seminar, University of Illinois, January 5, 1949

Design of Plant for Removal of H_2S from Natural Gas

Because of the poisoning effect of sulfur compounds on catalytic agents it is necessary to reduce the H_2S concentration from 0.002 moles per mole of gas to 1.6×10^{-6} moles per mole of gas. Further removal of sulfur is useless because at this lower concentration the rate of catalyst loss due to sulfur poisoning is far below the normal rate of deterioration.

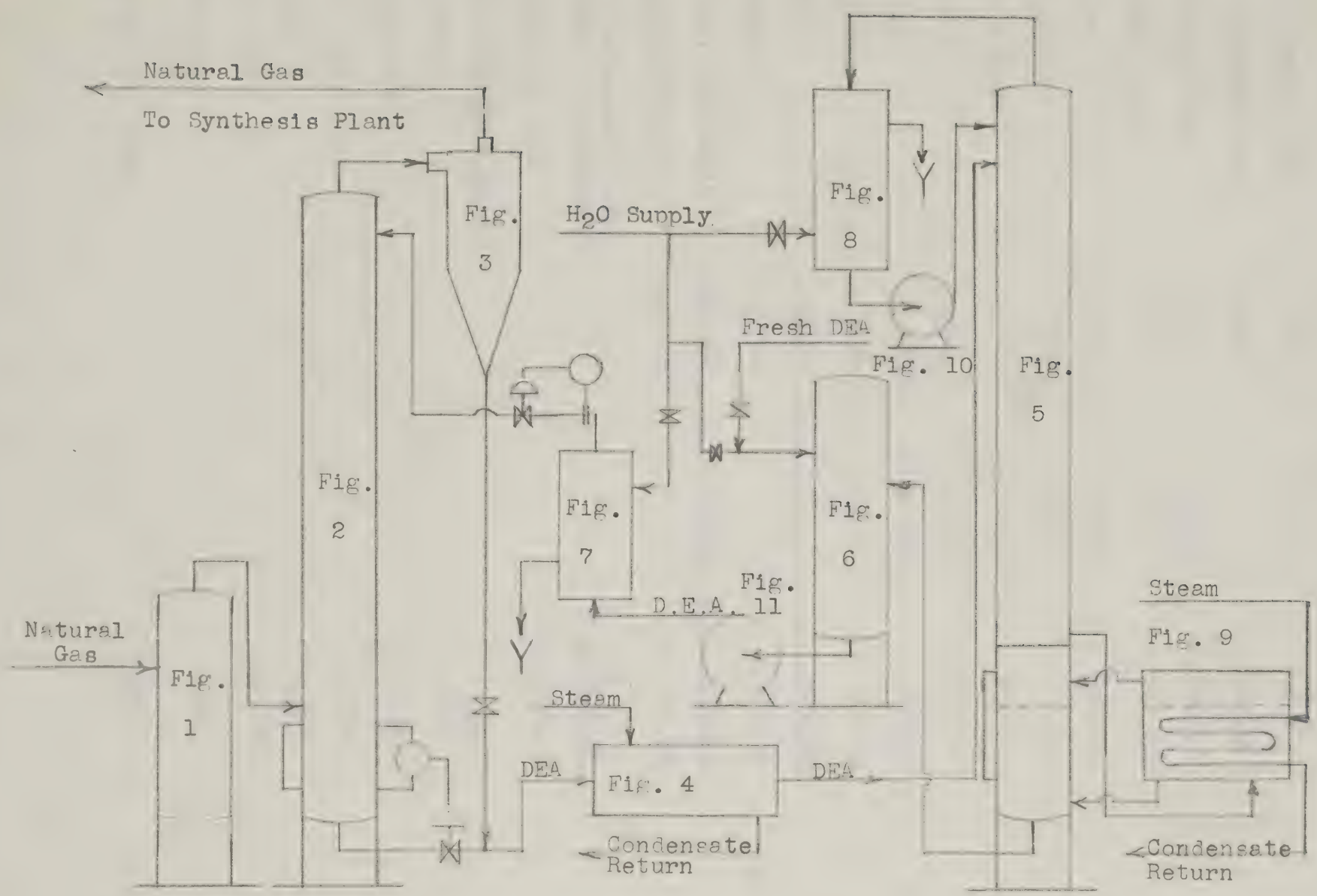
The H_2S is to be removed from 6.2×10^7 SCF/Day (2.57×10^6 SCF/hr.) by absorption in 50% diethanolamine solution (D.E.A.). This flow rate corresponds to 14.3 moles/hr. of H_2S or 11,660 Lb/Day.

The natural gas enters and leaves the surge tank (Fig. 1 Page 5) at 320 psia, 95°F, saturated with water vapor. From the surge tank it passes to the H_2S absorber (Fig. 2 Page 6) which operates isothermally at 95°F, 320 psia. The H_2S free gas leaves the tower at 315 psia and passes to the cyclone separator (Fig. 3 Page 7) where any entrained D.E.A. is removed. The gas then passes to the synthesis gas reactor.

The H_2S rich D.E.A. is heated in an exchanger (Fig. 4 Page 8) using 30 psig steam to 205°C. The hot solution is then throttled to low pressure and passes to the stripping tower (Fig. 5 Page 9) where the H_2S is stripped out of the solution by steam distillation. The lean D.E.A. solution leaves the bottom of the tower at 230°F and passes to the surge tank (Fig. 6 Page 10) where water and D.E.A. losses are made up. The solution is then pumped at 350 psia to the cooler (Fig. 7 Page 11) where it is cooled to 95°F before being returned to the absorption tower.

The vapor from the stripping tower is sent to a partial condenser (Fig. 8 Page 12) where 98% of the condensible vapor is removed. The effluent gas containing 90% H_2S and 10% H_2O is sent to the furnace for disposal.

A flow diagram for the H_2S removal plant is included (page 3). Design calculations for individual pieces are contained in the appendix (pages A-2 to A-44); cost calculations (pages A-45 to A-51). A summary of investment costs, yearly costs, and hourly operating costs is given on (page 4).



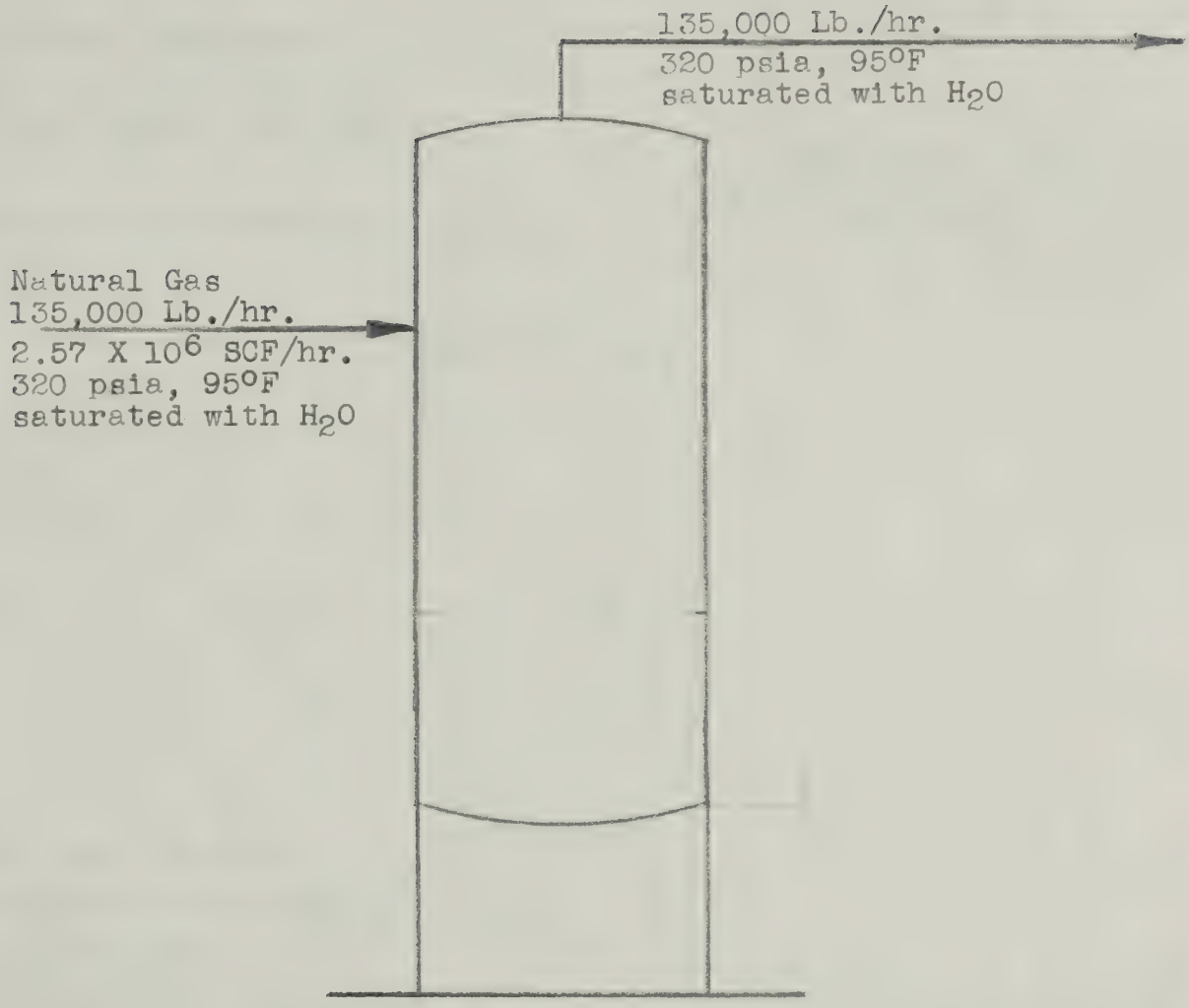
FLOW DIAGRAM — H_2S Removal

COST SUMMARY

UNIT	INVEST-MENT $	DEPRECIATION PER YEAR %	COST PER YEAR $
D. E. A. Surge Tank	940	5	47.00
Feed Gas Surge Tank	8,650	5	432.00
Absorption Tower	34,100	6-2/3	2,270.00
Stripping Tower	20,150	6-2/3	1,343.00
Reboiler	2,465	6-2/3	164.00
D.E.A. Heater	377	6-2/3	25.00
D.E.A. Cooler	1,722	6-2/3	115.00
Condenser	2,455	6-2/3	164.00
Main Pump	1,260	6-2/3	84.00
Reflux Pump	540	6-2/3	36.00
Cyclone	4,104	5	205.00
Basic Cost	76,763		4,885.00
Auxiliary Equipment	23,030	6-2/3	1,535.00
Constructed Cost	99,793		6,420.00
Yearly interest, Taxes, Insurance, & Overhead			11,976.00
TOTAL YEARLY COST			$18,396.00

Hourly Operating Cost

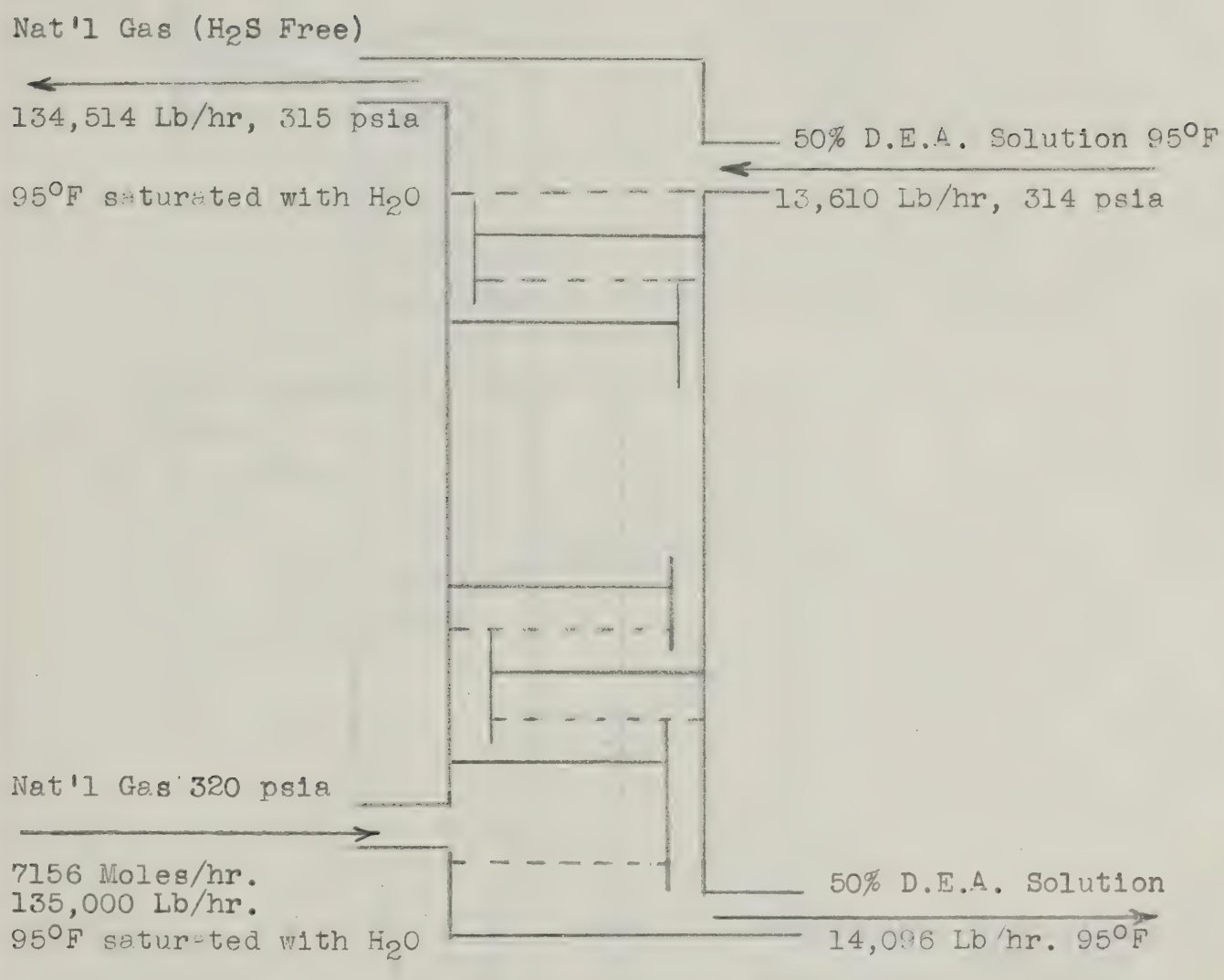
ITEM	Cost $/hr.
Cooling water	0.567
Power	0.059
Operators	2.000
D.E.A. (Make up)	0.250
Fixed charges	2.100
TOTAL HOURLY COST	4.976



Height	15 ft.
Diameter	10 ft.
Volume	1178 cu. ft.
Wall Thickness	1.35 in.
Weight	34,600 Lb.
Cost (installed)	$8,650.00

NATURAL GAS SURGE TANK

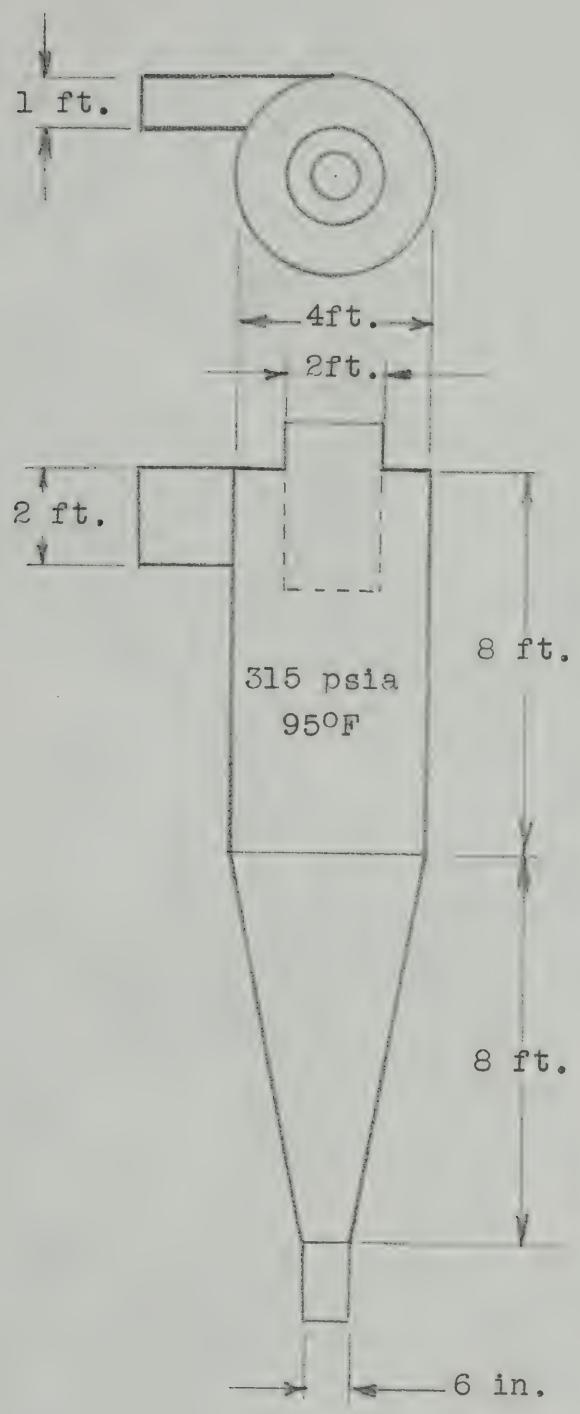
Fig. 1



Height	57 ft.
No. of Plates	27
Tray spacing	24 in.
Diameter	7 ft.
Wall Thickness	1 in.
Weight	58,300 lb.
Cost (installed)	$34,100.00

H_2S ABSORPTION TOWER

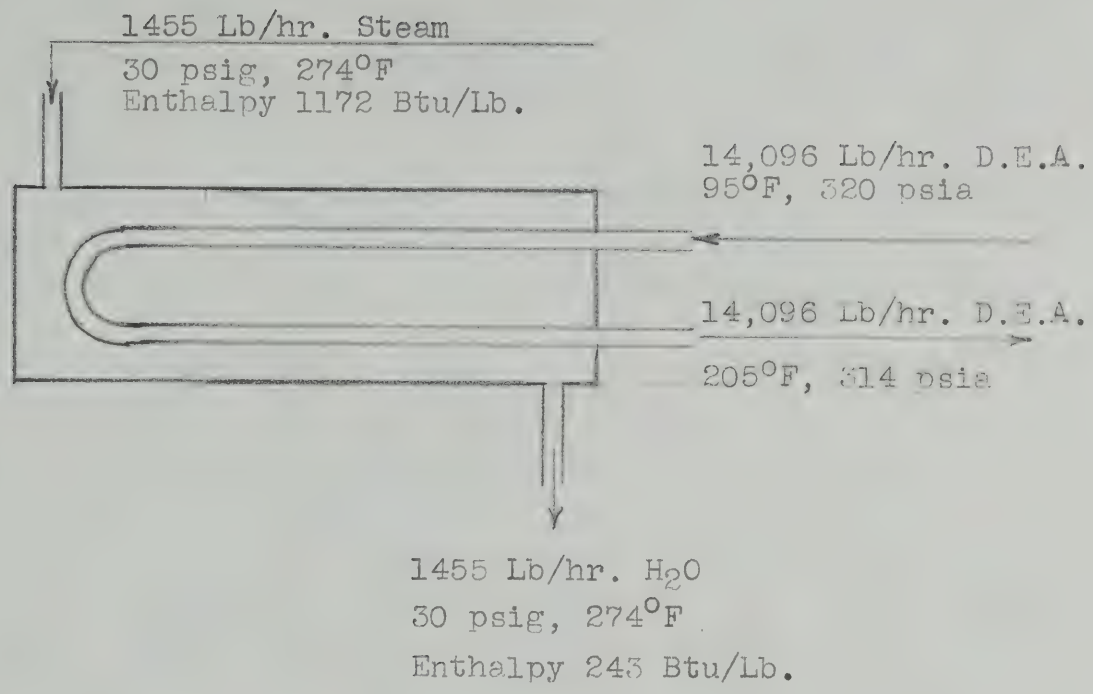
Fig. 2



Throughput--134,514Lb/hr.
Nat'l Gas
Minimum size droplets
removed--20 Microns
Pressure Drop--.3 psi
Wall Thickness--0.625 in.
Weight--5130 Lb.
Cost (installed)--$4,104

CYCLONE SEPARATOR

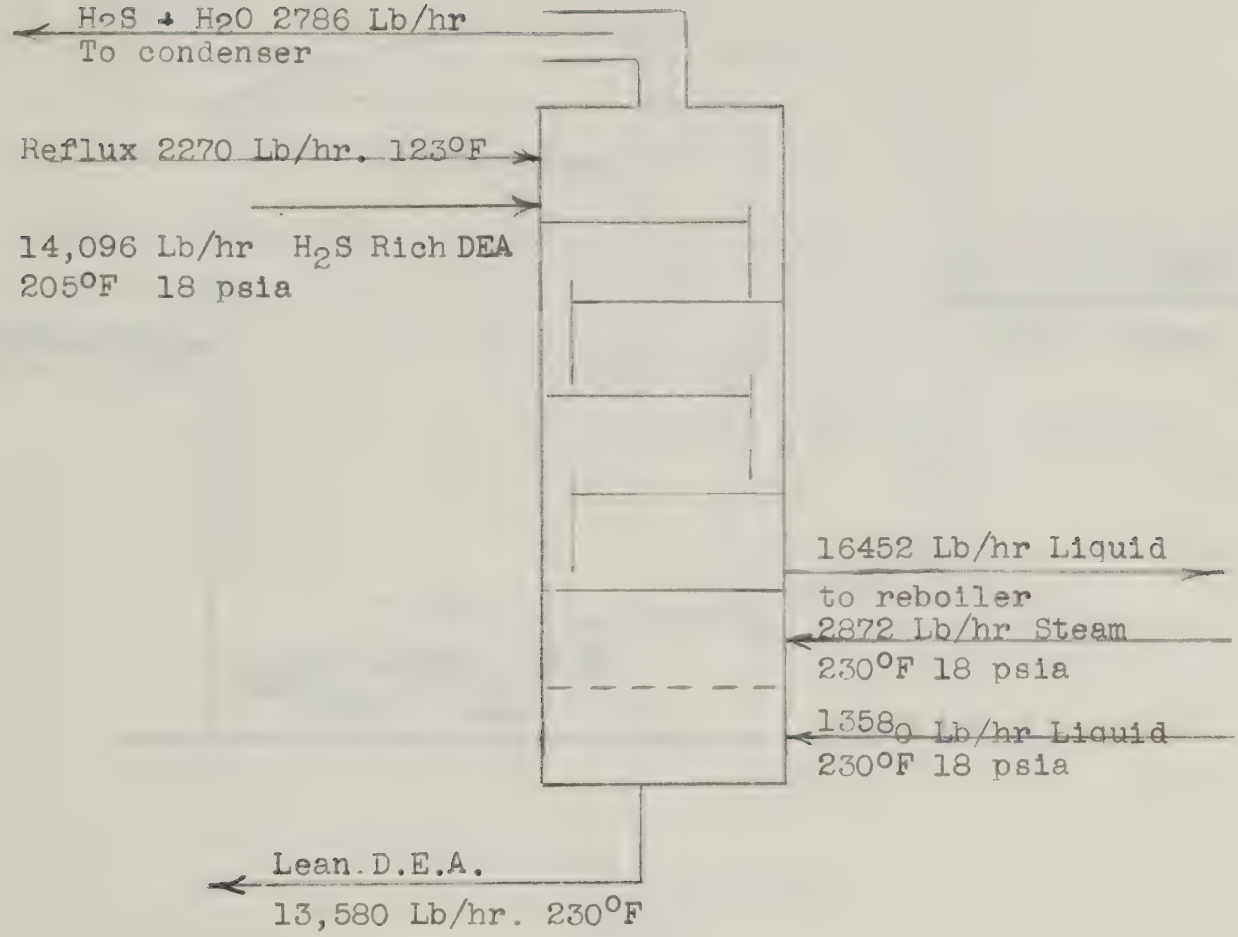
Fig. 3



Length	9 ft.
Tubes	20, ½ in. X 18 B.W.G.
Passes	2, 10 tubes per pass
Area	18.9 sq. ft. (inside)
Heat Load	1,350,000 Btu/hr.
Average Coefficient	693 Btu/hr.°F sq. ft. (inside)
Δt (mean)	115°F
Pressure Drop	5.5 psi.
Cost (installed)	$377.00

DIETHANOLAMINE HEATER

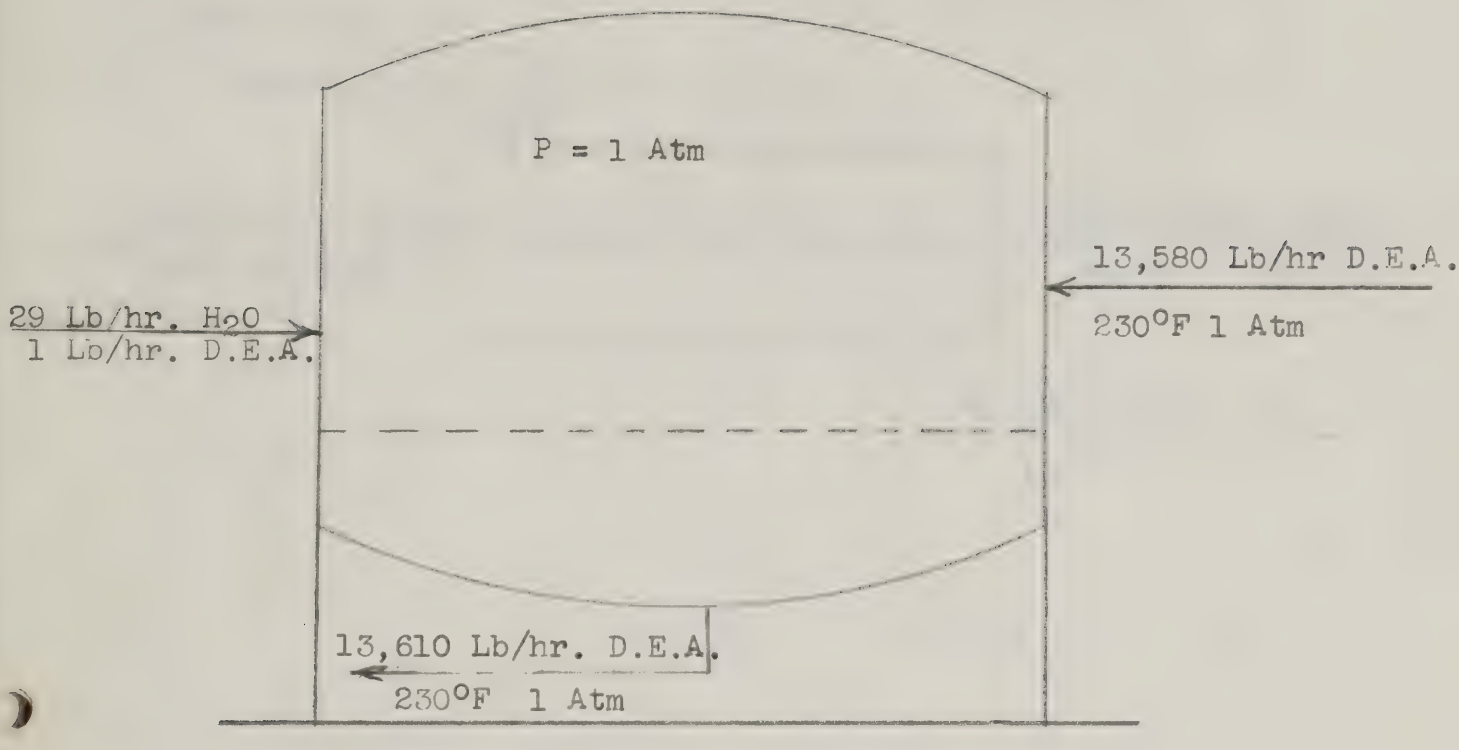
Fig. 4



Height	125 ft.
No. of Plates	27
Tray spacing	24 in.
Diameter	3 ft.
Weight	13,800 Lb.
Cost (installed)	$20,150.00

D.E.A. STRIPPING TOWER

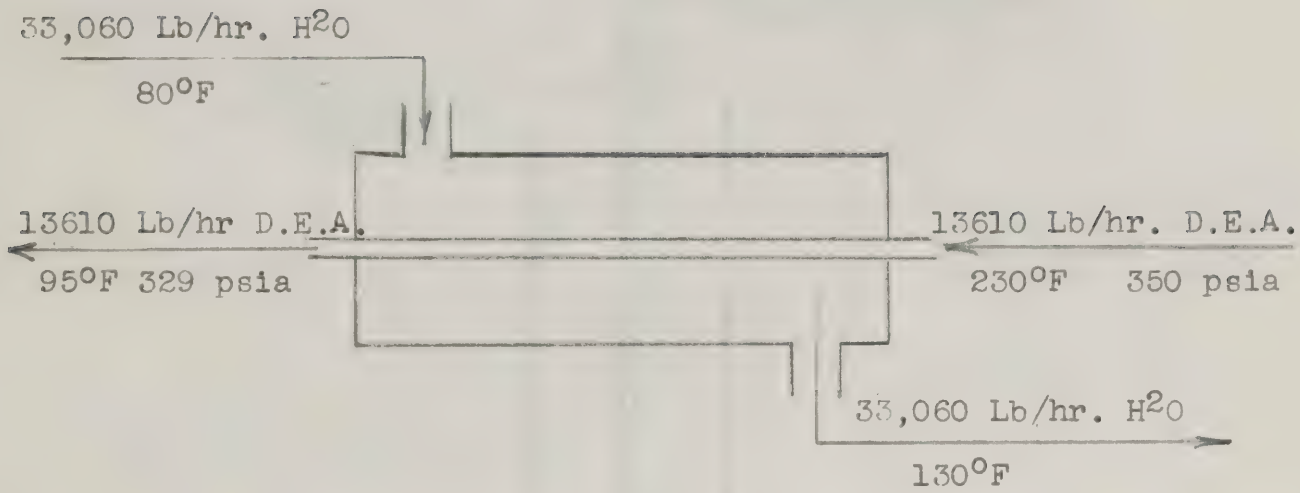
Fig. 5



Height	10 ft.
Diameter	15 ft.
Volume	1765 cu. ft.
Wall Thickness	0.125 in.
Weight	3750 lb.
Cost (installed)	$940.00

DIETHANOLAMINE SURGE TANK

Fig. 6

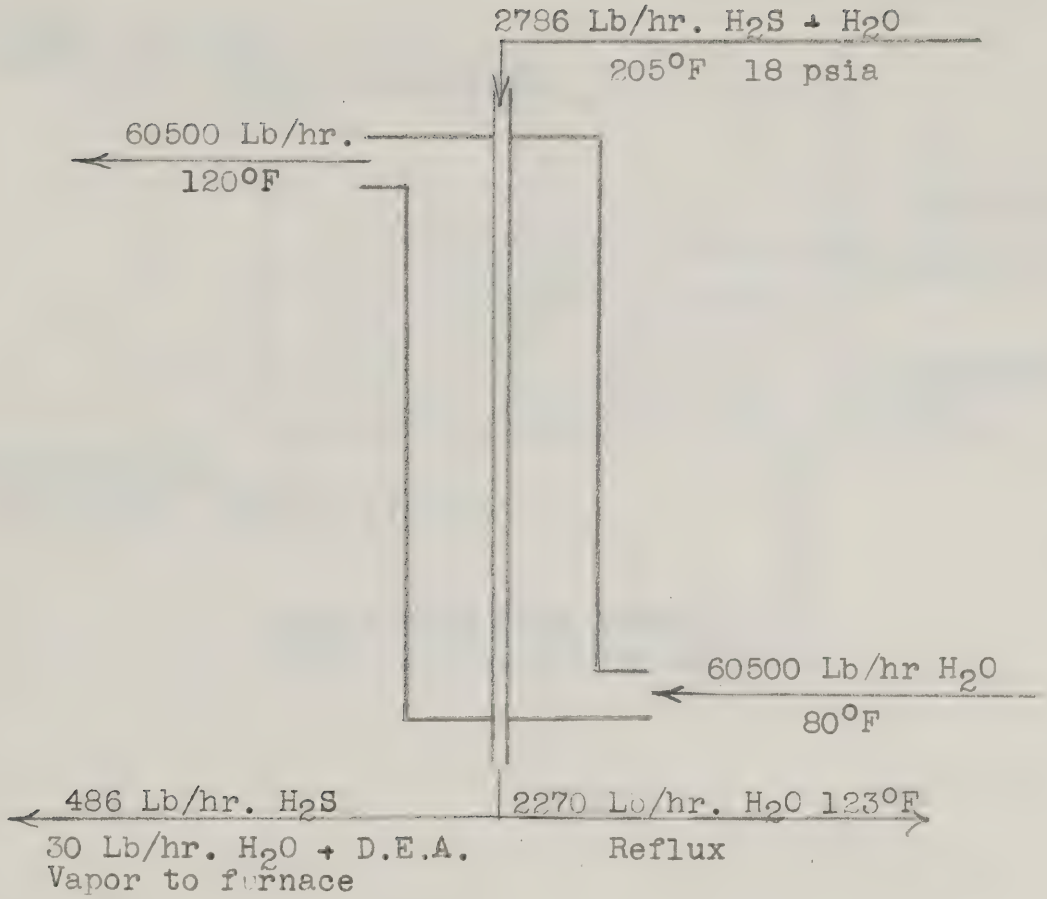


Five Units Required in series

Length	15 ft./unit
Tubes	10 - ½" X 18 B.W.G.
Baffle Spacing	10 in.
Total Area	78.9 sq. ft.
Total Heat Load	1,653,000 Btu/hr.
Average Coefficient	487 Btu/hr.°F sq. ft. (inside)
Δt mean	44.8°F
Overall Pressure drop	21 psi
Cost (installed)	$1,722.00

DIETHANOLAMINE COOLER

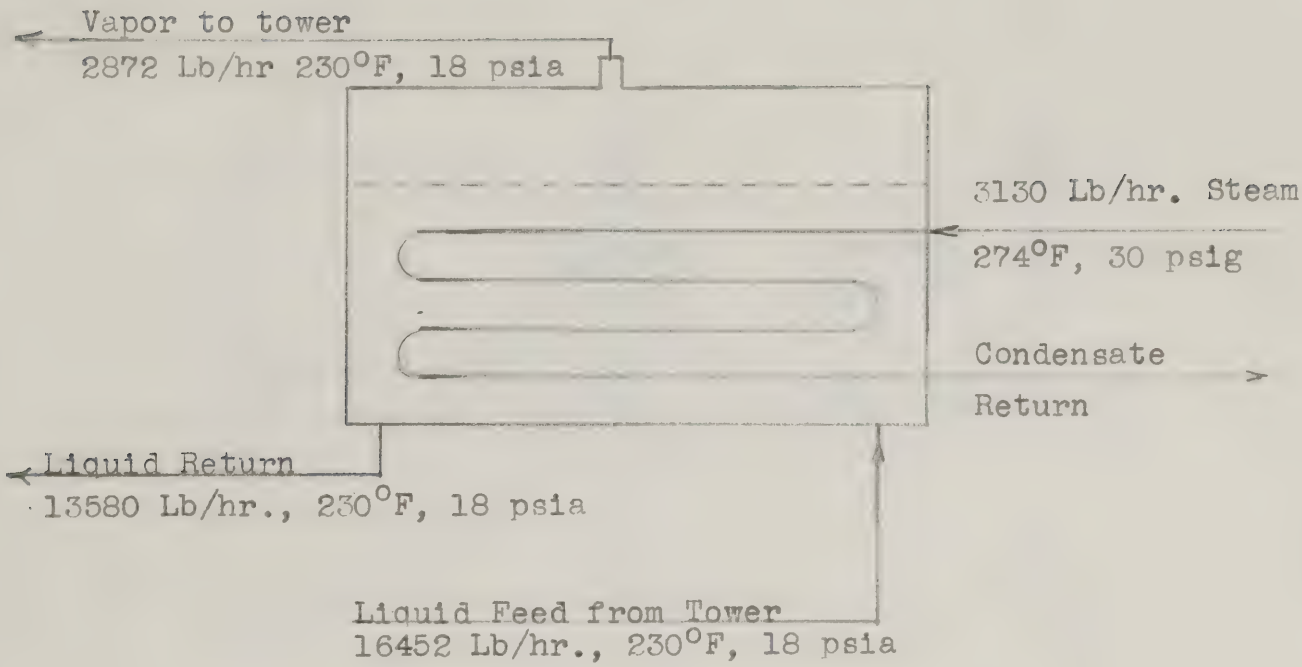
Fig. 7



Two units in series required	
Length/unit	36 ft.
Tubes	19 - ¼" X 18 B.W.G.
Baffle Spacing	8 in.
Total Area	131 sq. ft.
Total Heat Load	2,420,000 Btu/hr.
Average Coefficient	300
Δt mean	61.6°F
Cost (installed)	$2,455.00

STRIPPING TOWER VAPOR CONDENSER

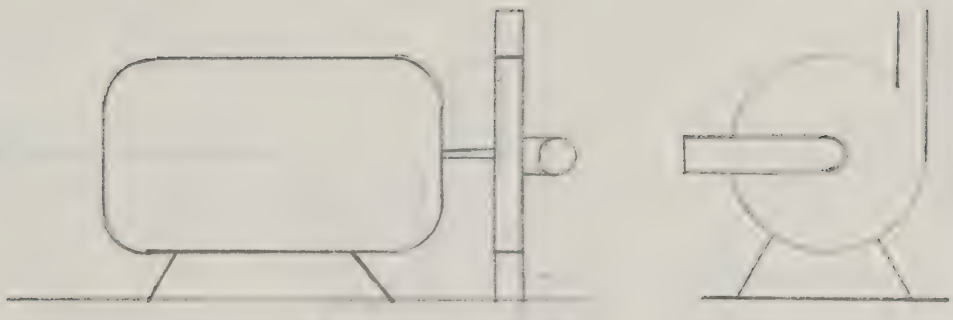
Fig. 8



Tubes	1"OD X 14 BWG
Area	100 sq. ft.
Heat Load	2,765,000 Btu/hr.
Heat Transfer Coefficient	1168 Btu/hr °F ft.^{2}
Δt	30°F
Cost (installed)	$2,465

STRIPPING TOWER REBOILER

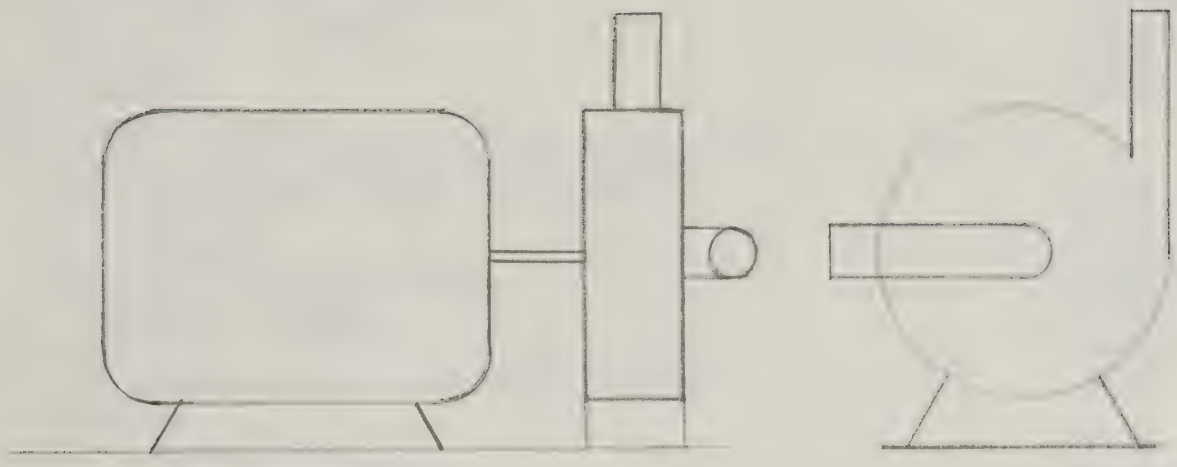
Fig. 9



Throughput	2,270 Lb/hr.
Pressure increase	50psi
Horse power	1.3
Power consumption	1.31KW
Cost (installed)	
Includes motor & standby unit	$540

REFLUX PUMP

Fig. 10



Throughput	13,610 Lb/hr.
Pressure increase	335 psi
Horse power	10.6
Power Consumption	8.9KW
Cost (installed)	
Includes motor & stand by unit	$1,256

MAIN PUMP

Fig. 11

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H_2S REMOVAL SYSTEM

OVERALL H_2S BALANCE

Maximum Allowable Sulfur $= \frac{0.1 \text{ grains}}{100 \text{ S.C.F.}}$

$$\frac{0.1}{7000 \times 32} = 4.464 \times 10^{-7} \frac{\text{Lb. Atoms Sulfur}}{100 \text{ S.C.F.}}$$

$$= 4.464 \times 10^{-7} \frac{\text{Lb. Moles } H_2S}{100 \text{ S.C.F.}}$$

$$4.464 \times 10^{-7} \times 0.729 \times 4.92 = 1.601 \times 10^{-4} \frac{ft^3 \, H_2S}{100 \text{ S.C.F.}}$$

final concentration $y_2 = 1.6 \times 10^{-6} \frac{\text{moles } H_2S}{\text{moles gas}}$

initial concentration $y_1 = 0.002 \frac{\text{moles } H_2S}{\text{mole gas}}$

H_2S Removed $= 0.002 - 1.6 \times 10^{-6} = 1.9984 \times 10^{-3}$ moles/mole gas

gas treated $= 6.16 \times 10^{7} \frac{ft^3}{DAY} = 2.567 \times 10^{6} \frac{ft^3}{hr.}$

moles gas treated/hr. $= \frac{2.567 \times 10^{6}}{0.729 \times 492} = 7.156 \times 10^{3} \frac{\text{moles}}{hr.}$

H_2S Removed $= 7.156 \times 10^{3} \times 1.9984 \times 10^{-3} = 14.3 \frac{\text{moles}}{hr.}$

Equilibrium Data for H_2S, 50% D.E.A. System

Reference: Perry, Page 1133, Fig. 4

Data are plotted P_{H_2S} vs $\frac{ft.^3\ H_2S\ (60°, Atm)}{ft.^3\ D.E.A.}$

Conversion to mole fraction:

Basis: 100 Lb. 50% D.E.A.

M.W. D.E.A. = 105.1

Moles D.E.A. = $\frac{50}{105.1}$ = 0.476

Moles H_2O = $\frac{50}{18}$ = 2.777

Total = 3.253

MW_{av} (50% D.E.A.) = $\frac{100}{3.253}$ = 30.74 Lb./Lb mole

Assume: Volumes are additive

ρ_{DEA} = 1.0975 × 62.4 = 68.45 Lb/ft³

ρ_{H_2O} = 62.4 Lb/ft³

V_{DEA} = $\frac{50}{68.45}$ = 0.730 ft³

V_{H_2O} = $\frac{50}{62.4}$ = 0.801 ft³

V_{Total} = = 1.531 ft³

$\rho_{50\%\ D.E.A.}$ = $\frac{100}{1.531}$ = 65.35 Lb./ft.³

$\frac{\text{Moles solution}}{ft^3}$ = $\frac{3.253}{1.531}$ = 2.125

$$\frac{\text{Moles } H_2S}{ft^3\ H_2S\ (60°, 1\ Atm)} = \frac{1}{0.729 \times 520} = \frac{0.00264 \text{ Moles } (H_2S)}{ft^3\ H_2S}$$

$$\frac{\text{Moles } H_2S}{\text{Moles D.E.A.}} = X = \frac{0.00264}{2.125} \times \frac{ft^3\ (H_2S)}{ft^3\ \text{D.E.A.}} = 0.001242 \times \frac{ft^3\ H_2S}{ft.^3\ \text{D.E.A.}}$$

$$x = \frac{X}{1+X} = \frac{0.001242}{1.001242} \frac{ft^3\ H_2S}{ft^3\ \text{D.E.A.}} = 0.00124 \frac{ft^3\ H_2S}{ft^3\ \text{D.E.A.}}$$

ABSORBER DESIGN

Tower will operate at 95°F 310 psig ≈ 325 psia

$$\frac{325 \times 760}{14.7} = 16{,}800 \text{ mm Hg}$$

$$y = \frac{P_{H_2S}}{16{,}800}$$

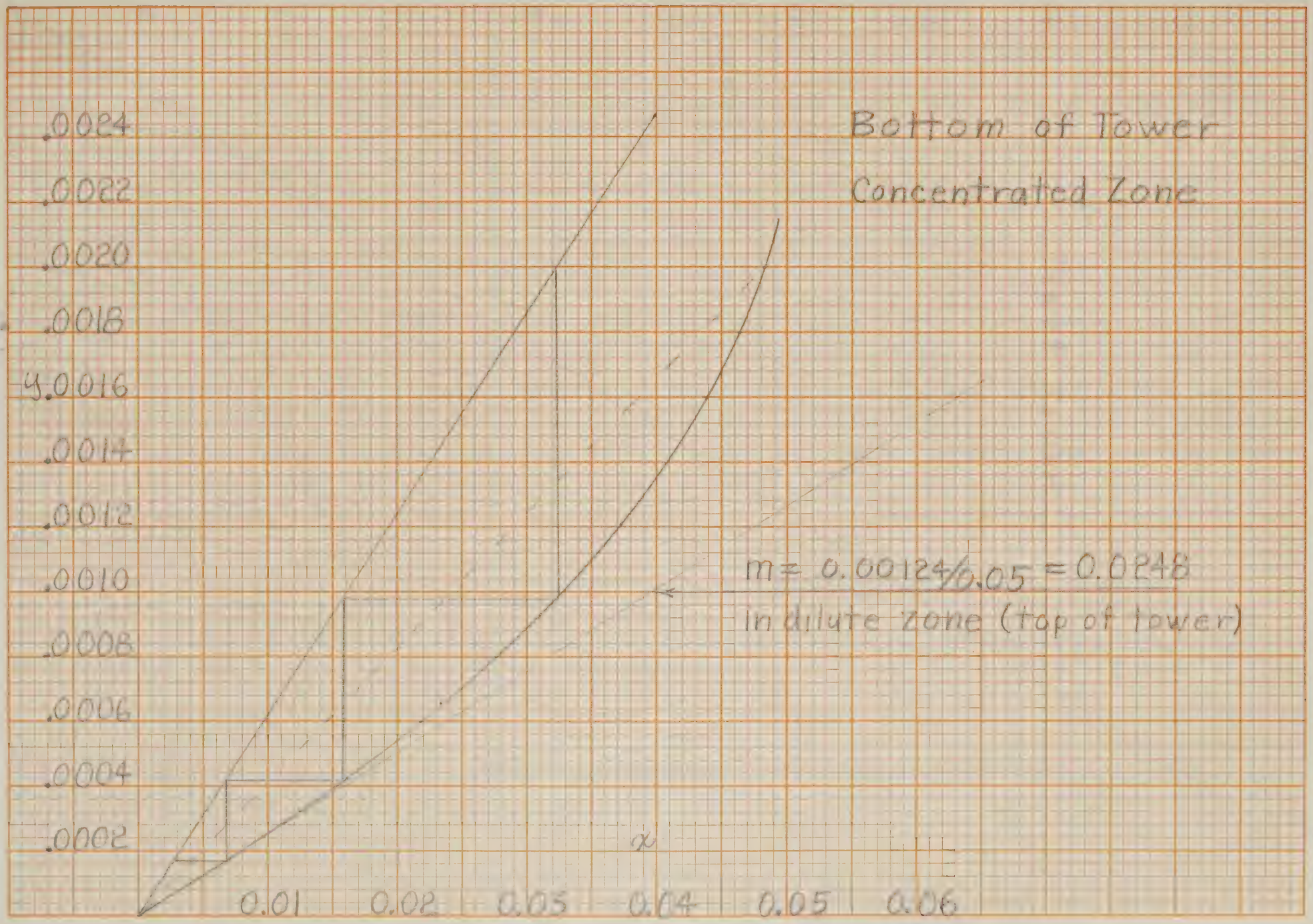
Data:

ft^3 H_2S (60° 1 Atm) / ft^3 D.E.A.	10	20	30	40	50	60	70	80
x Moles H_2S / Moles Solution	0.0124	0.0248	0.0372	0.0496	0.0620	0.0744	0.0868	0.0992
P_{H_2S} (mm Hg)	5	12	20	35	60	85	135	230
y Moles H_2S $\times 10^3$ / Mole gas	0.307	0.737	1.228	2.15	3.68	5.22	8.29	14.12

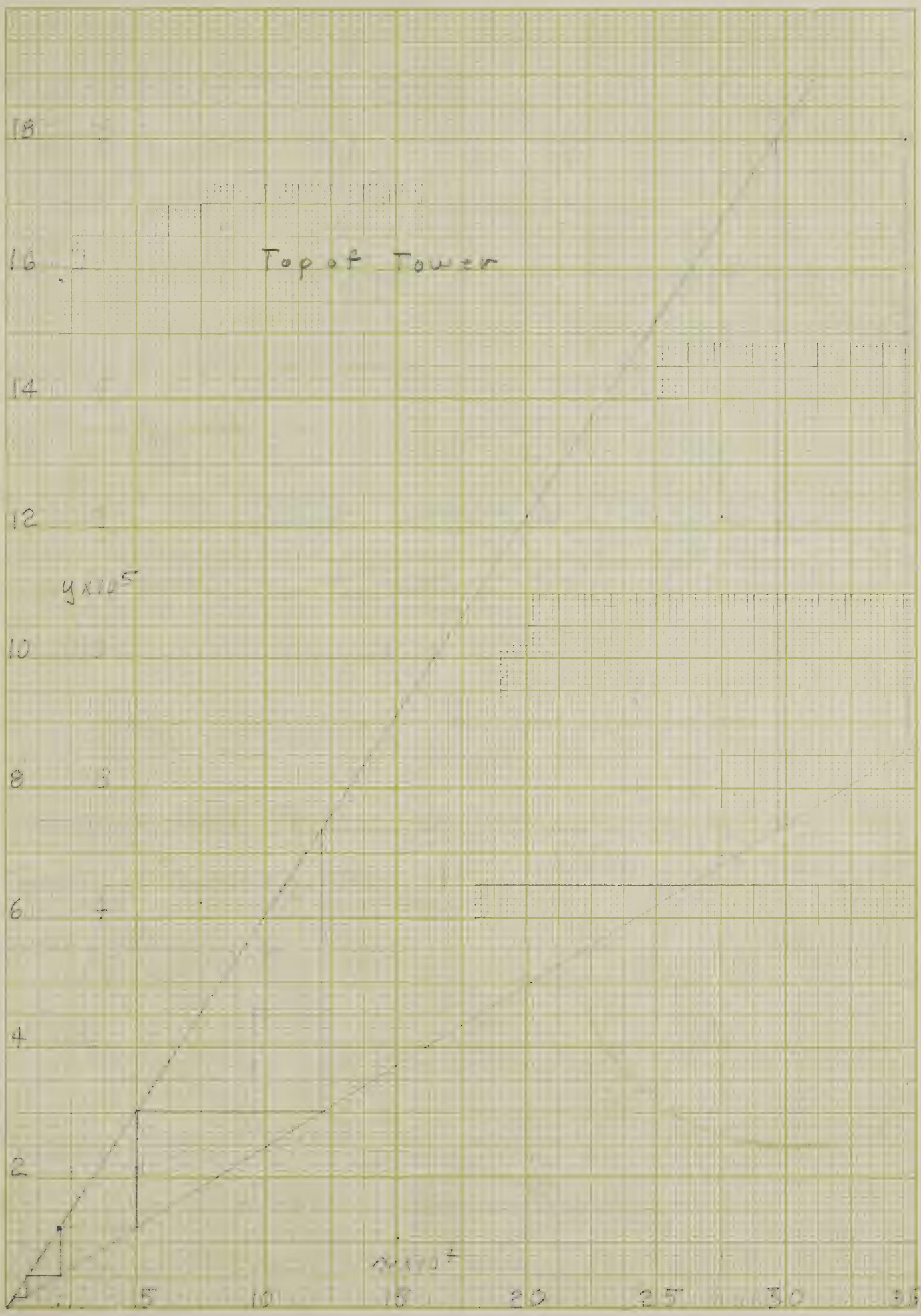
from x, y diagram L/V (minimum) $= \frac{0.00165}{0.04} = 0.04125$

Use $L/V = 1.5\ L/V$ (minimum) $= 1.5 \times 0.04125 = 0.06187$

KEUFFEL & ESSER CO., N. Y. NO. 359-6
5 × 5 to the ½ inch.



A-6


18
16
14
12
10
8
6
4
2
Top of Tower
y x10^5
5
10
15
20
25
30

Overall Material Balance on Tower

$$y_1 - y_2 = \frac{L}{V}(x_1 - x_2)$$

$$0.002 - 0.0000016 = 0.06187\,(0.032326 - x_2)$$

$$x_2 = 0.032326 - \frac{0.0019984}{0.06187} = 0.032326 - 0.03230 = 0.000026$$

Material Balance on Bottom Section

$$0.002 - 0.00018 = 0.00182 = 0.06187\,(0.0325 - x')$$

$$x' = 0.032326 - \frac{0.00182}{0.06187} = 0.032326 - 0.029416 = 0.00291$$

Material Balance on Top Section

$$0.00018 - 0.0000016 = 0.06187\,(0.00291 - x_2)$$

$$x_2 = 0.00291 - \frac{0.0001784}{0.06187} = 0.00291 - 0.002883 = 0.000027$$

Checks Overall Balance

Theoretical Plates Required

Bottom of Tower $N_{P_1} = 3$
Top of Tower $N_{P_2} = 5$
Total $N_P = 8$
Assuming Tray efficiency of 30%,
Tower requires $\frac{8}{0.30} = 26.7$ Plates

Use 27 Plate Tower

$\dot{V} = 7.156 \times 10^3$ Moles/hr. (See page ___)

$L = 7.156 \times 10^3 \times 0.06187 = 442.8$ Moles/hr.

$L = 442.8 \times 30.74 = 13{,}610$ Lb/hr.

Extension of Equilibrium Data

Data are plotted as log P_{H_2S} vs $\frac{1}{T}$ at constant x values and extrapolated to 0.00145 (Page).

$$\left(0.00145 = \frac{1}{690} = \frac{1}{230+460} \right)$$

By following the constant $\frac{1}{T}$ lines, data are obtained for P_{H_2S} vs x at $T = 205°F$ and $T = 230°F$. These data are plotted (page). Allowance is made for temperature gradient, through stripper, by joining the two curves with a smooth curve tangent to each curve of the appropriate end conditions:

Tower Top: $x = 0.03233$
$T = 205°F$

Tower Bottom: $x = 0.000026$
$T = 230°F$

From this tangent curve are obtained the Pressure concentration Data.

P_{H_2O} (205°F) = 661 mm Hg

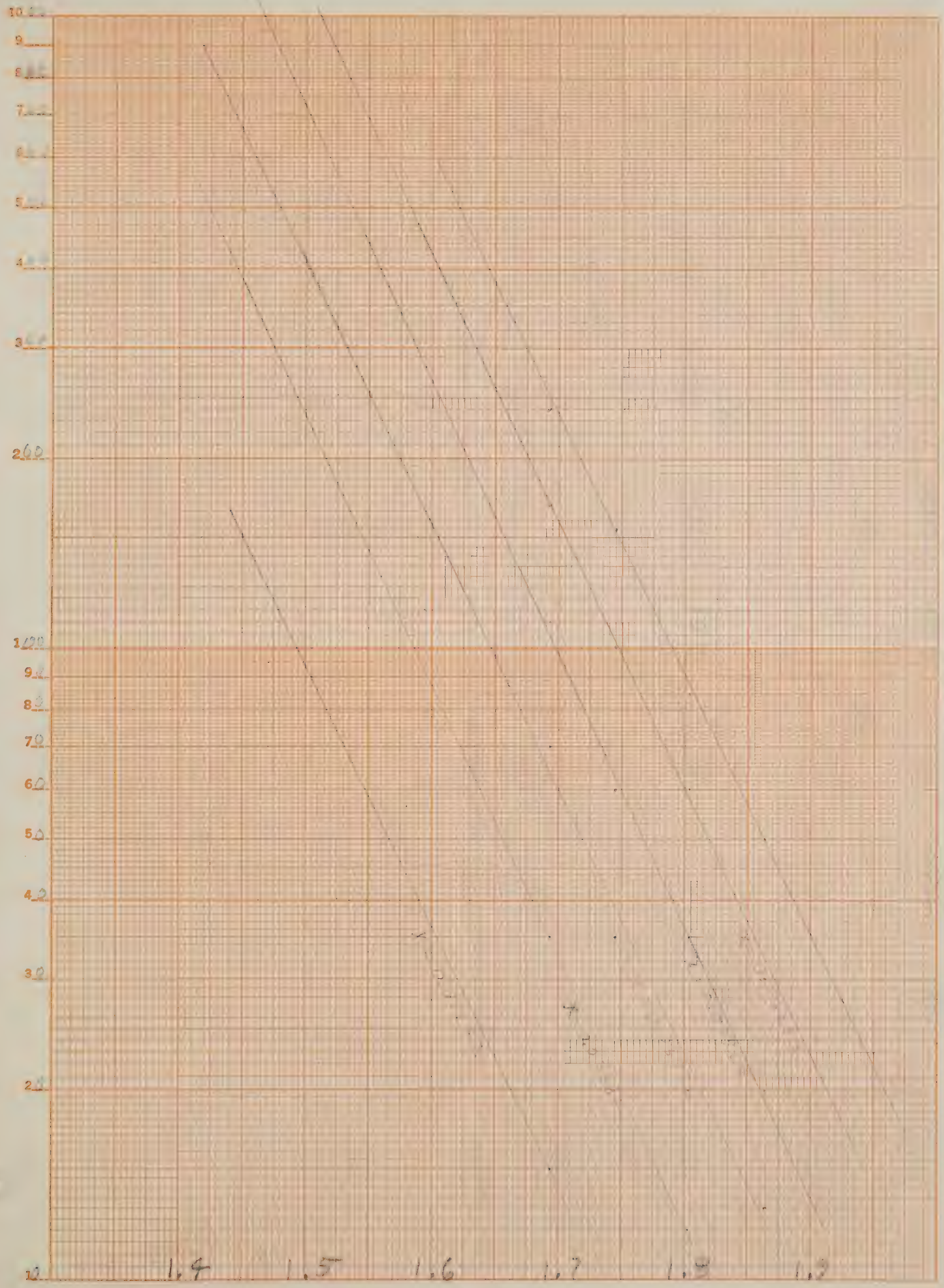
P_{H_2S} (205°F, $x = 0.0323$) = 280 mm Hg

Total Pressure = 280 + 661 = 941 mm Hg.

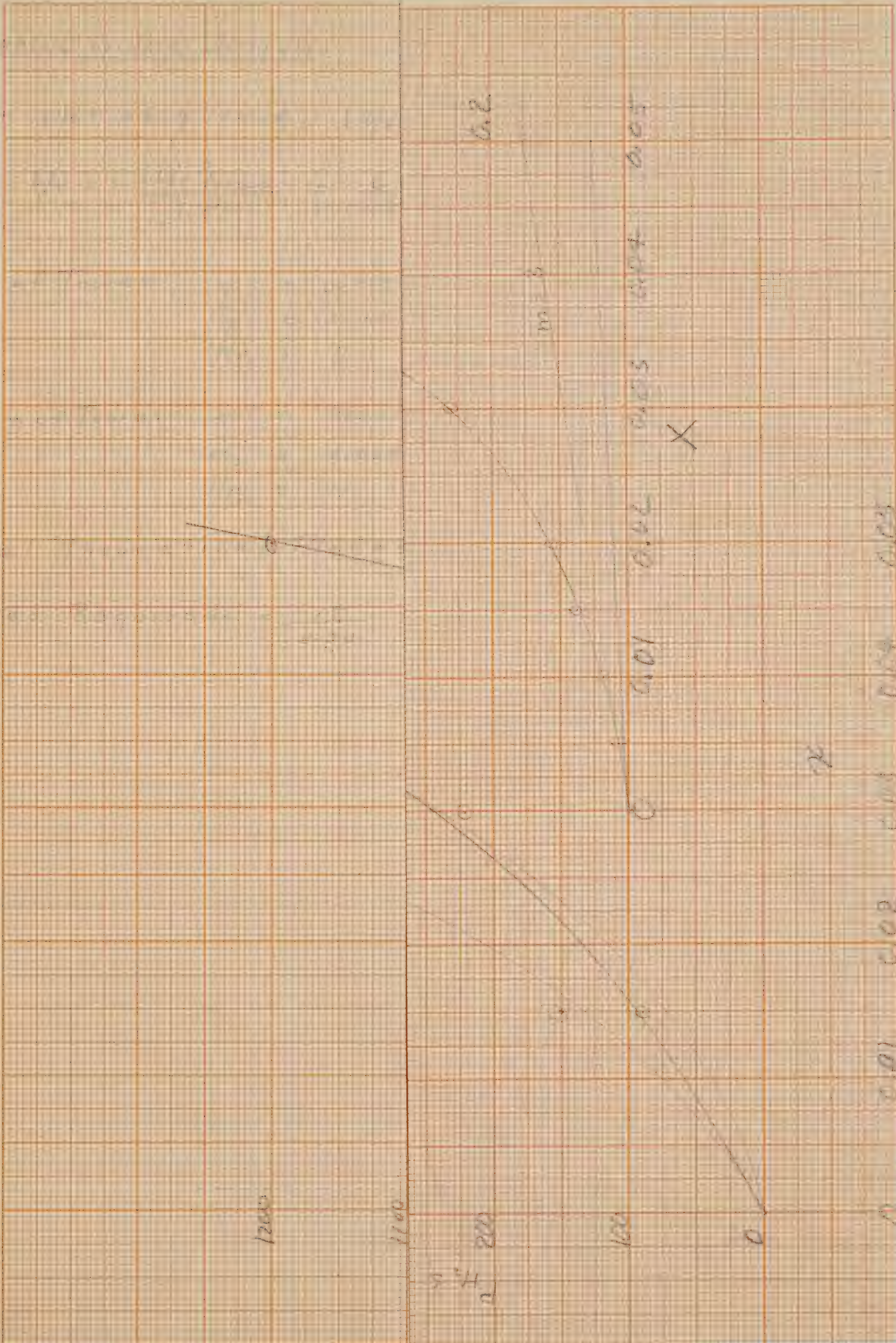
$$Y = \frac{P_{H_2S}}{941}$$

Data from cross plot

x	P_{H_2S} / mmHg	y
0.0025	15	0.01594
0.0050	29	0.03082
0.0075	44	0.04676
0.0100	59	0.0627
0.0150	94	0.0999
0.0200	135	0.1434
0.0250	184	0.1955
0.0300	245	0.2604
0.0350	323	0.3432



KEUFFEL & ESSER CO., N. Y. NO. 359-63
Semi-Logarithmic, 2 cycles × 10 to the ½ inch, 5th lines accented
MADE IN U. S. A.



Y
X
1.6
1.4
1.2
1.0
0.8
0.6
0.4
0.2
0.05
0.04
0.03
0.02
0.01
0.01
0.02
0.03
0.04
0.05
1200
1100
1000
900
800
700
600
500
400
300
200
100
0

STRIPPING TOWER DESIGN

From plot of y vs x, $(L/V)_{max} = \frac{0.18}{0.030} = 6.00$

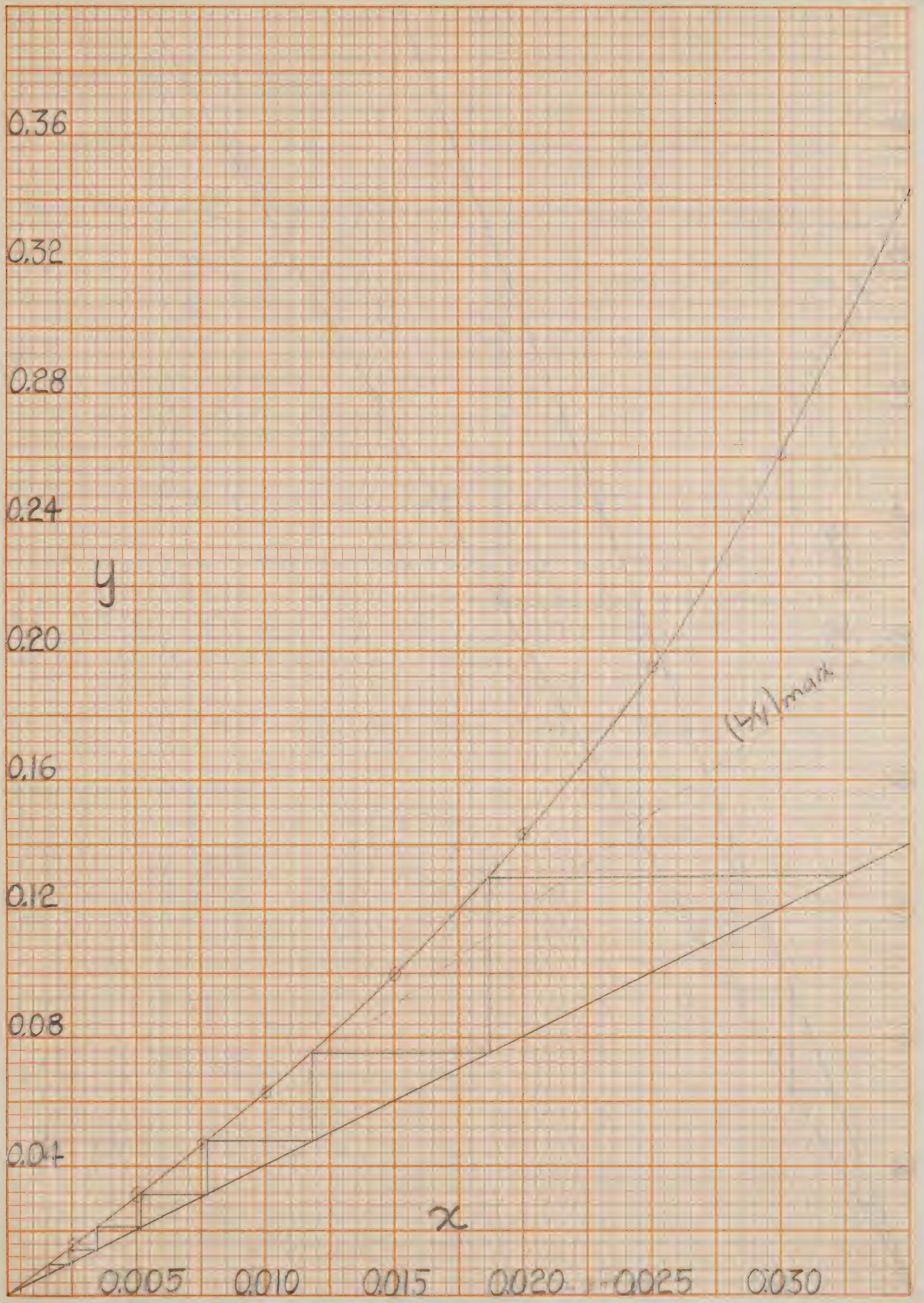
Use $L/V = \frac{(L/V)_{max}}{1.5} = \frac{6}{1.5} = 4.00$

Top of Tower $x_1 = 0.0323$
$x' = 0.0016$
$N_{P_1} = 7$

Bottom of Tower $x' = 0.0016$
$x_2 = 0.000025$
$N_{P_2} = 11$

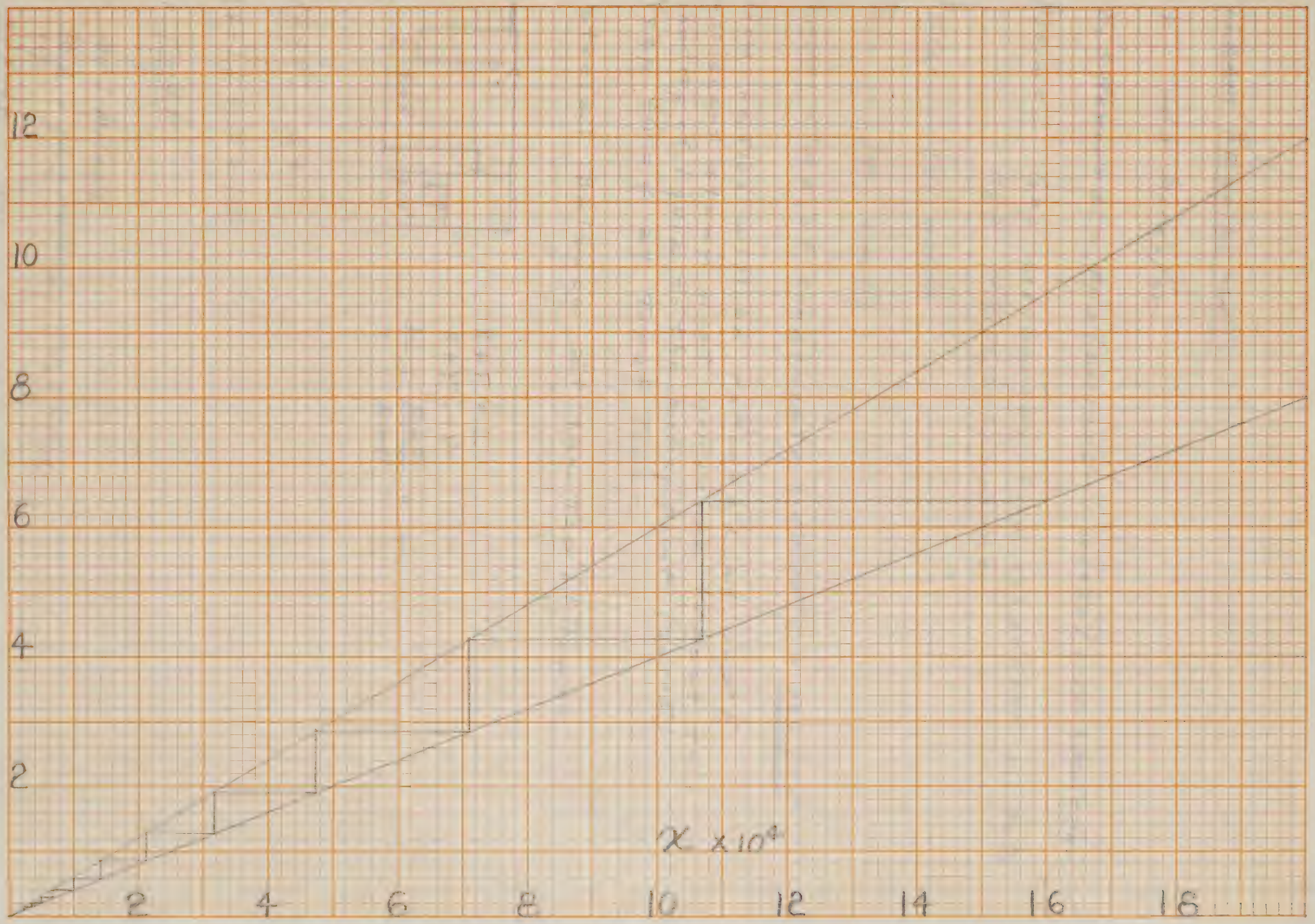
Total theoretical Plates $= 7 + 11 = 18$

Plates Required $= \frac{18}{0.30} = 60$ Actual Plates

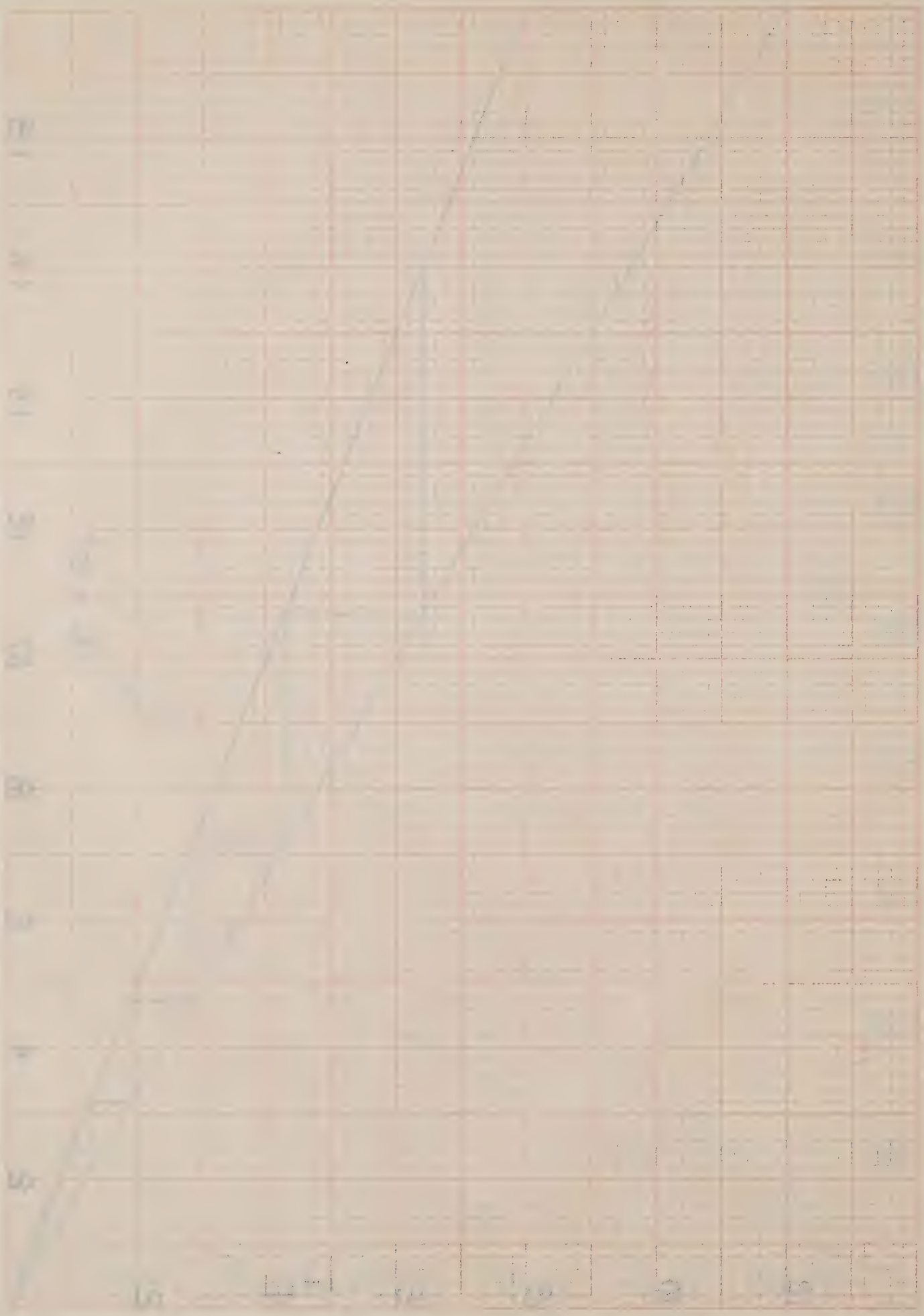


KEUFFEL & ESSER CO., N. Y. NO. 359-6
5 × 5 to the ½ inch.
MADE IN U. S. A.

KEUFFEL & ESSER CO., N. Y. NO. 359-6
5 X 5 to the ½ inch
MADE IN U. S. A.



A 15



STEAM REQUIREMENTS

$L = 443$ Moles/hr.

H_2S Removed $= 443(0.032326 - 0.000026) = 14.3$ Moles/hr.

Assume: Gas discharged from condenser is 90% H_2S 10% H_2O by volume

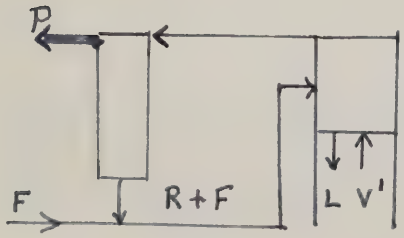
$P_{Total} = 941$ mm Hg

$P_{H_2S} = 94.1$ mm Hg

Temperature of condensate $= 50.34°C = \underline{\underline{122.6°F}}$

Since reflux will be returned at 123°F, sufficient steam must condense to heat this quantity of liquid to the boiling point (205°F)
Keep $L/V = 4$ at Top of Tower.

Material Balance at the Top of Tower.



$$R + F = L$$
$$R + P = V$$
$$\frac{L}{V} = \frac{R+F}{R+P} = 4$$

$$4(R+P) = R+F$$

$$R(4-1) = F - 4P$$

$$R = \frac{F-4P}{3}$$

$F = 443$ Moles/hr.

$P = \frac{14.3}{0.9} = 15.9$ Moles/hr. $(H_2S + H_2O)$

$R = \frac{443-64}{3} = \frac{379}{3} = 126$ Moles/hr.

Sensible Heat to raise Reflux from 123 to 205°F

$Q_1 = 126 \times 18 \times 82 = 185,980$ Btu/hr

Steam condensed $= \frac{185,980}{960 \times 18} = 10.8$ Moles/hr

Sensible Heat change through tower

$Q_2 = 25°F \left[137 \times 18 + 443 \times 30.7 \times 0.9 \frac{Btu}{Lb.°F} \right]$

$Q_2 = 25 \left[14,200 \right] = 367,500$ Btu/hr

Steam condensed $= \frac{367,500}{960 \times 18} = 21.3$ Moles/hr.

Steam overhead $= V - 14.3 = 126 + 15.9 - 14.3 = 127.6$ Moles/hr

Total Steam Produced in Reboiler

$10.8 + 21.3 + 127.6 =$ 159.7 Moles/hr

$\frac{L}{V}$ at bottom $= \frac{443 + 126 + 32.1}{159.7} = 3.76$

$Q_R = 160 \times 18 \times 960 = 2,765,000$ Btu/hr.

USE $\Delta_m = 40°$

Heating Steam (Thermal Efficiency = 0.95)

$T = 230 + 40 = 270°F$
$P = 42$ psia $= 27$ psig
$\Delta H_v = 930$ Btu/Lb.

Steam Required $= \frac{2,765,000}{930 \times 0.95} =$ 3,130 Lb/hr

WATER and D.E.A. LOSSES

P_{DEA} (123°F) = 1.7 mm Hg

P_{Total} = 941

Mole Ratio $\frac{DEA}{H_2S} = \frac{1.7}{939} = \frac{0.0018 \text{ Moles DEA}}{\text{Mole } H_2S}$

D.E.A. Lost = 15.9 × 0.0018 × 30.74 = 0.88 Lb/hr.

H_2O Lost = 15.9 − 14.3 = 1.6 Moles = 28.8 Lb/hr.

TOWER DIAMETERS

Basis: 6.2×10^7 ft^3 Natural Gas/Day

Absorber

$Q = 6.2 \times 10^7$ SCF/Day

Tray spacing = 24", Brown Tray spacing factor $B = 0.82$

Pressure = 300 psig, Barton pressure factor $C = 1.56$

ρ_g = S.G. of gas $= \frac{M.W}{29} = \frac{18.84}{29} = 0.65$ *

$T = 95°F = 555°R$

K = S.G. of Liquid $= \frac{65.35}{62.4} = 1.047$

$P = 315$ psia

$$A = \frac{\pi D^2}{4} = \frac{BQ}{10^6 C}\sqrt{\frac{\rho_g}{K}\frac{T}{P}} = \frac{0.82 \times 6.2 \times 10^7}{1.56 \times 10^6}\sqrt{\frac{0.65 \times 555}{1.047 \times 315}}$$

$$A = 32.6\sqrt{1.094} = 34.1 \text{ ft}^2$$

$$D = \sqrt{\frac{4 \times 34.1}{\pi}} = \sqrt{43.4} = 6.6 \text{ ft.}$$

USE 7 ft. COLUMN

Stripper

$Q = 24 \times 160 \times 0.729 \times 492 = 1.38 \times 10^6$ SCF/DAY

$P = \frac{941}{760} \times 14.7 = 18.2$ psia $= 3.5$ psig

$B = 0.82$, $T = 690°R$

$C = 0.78$ $\rho_g = \frac{18}{29} = 0.62$

$K = 1.047$

$$A = \frac{0.82 \times 1.38}{0.78} \sqrt{\frac{0.62 \times 690}{1.047 \times 18.9}} = 1.45 \sqrt{21.6}$$

$$A = 6.74 \text{ ft}^2$$

$$D = \sqrt{\frac{4 \times 6.74}{\pi}} = 2.93 \text{ ft}$$

USE 3 foot COLUMN

* Average M.W.

COMP.	Fraction	M.W.	Wt.
CH_4	0.798	16	12.77
C_2H_6	0.195	30	5.85
N_2	0.005	28	0.14
H_2S	0.002	34	0.07
	MW_{AV} =		18.83

REBOILER DESIGN

from McAdams Page 309

h (water, boiling at 212°F, ΔT = 30°F) = 8000 Btu/hr ft² °F

USE h (50% D.E.A., boiling at 230°F, ΔT = 30°F) = 4000 Btu/hr ft² °F

h_{steam} = 2500 Btu/hr ft² °F

Using 1" OD X 14 BWG Tubes

$$h_{Tube} = \frac{60 \times 12}{0.083} = 8670 \text{ Btu/hr ft}^2\,°F$$

$$\frac{1}{u} = \frac{1}{4000} + \frac{1 \times 1.000}{2500 \times 0.834} + \frac{1 \times 1.000}{8670 \times 0.917}$$

$$\frac{1}{u} = 0.00025 + 0.00048 + 0.000126 = 0.000856$$

u = 1168 Btu/hr ft² °F

ΔT = 30°

Q = 2,765,000 Btu/hr

$$A = \frac{2,765,000}{1168 \times 30} = 78.9 \text{ FT}^2$$

USE 100 ft² to Allow for Fouling

HEAT EXCHANGER DESIGN

C_p D.E.A. By Kopps Rule, $C_4 H_{11} O_2 N$ MW = 105

4C	–	4 × 2.8	=	11.2
11 H	–	11 × 4.3	=	47.3
2 O	–	2 × 6.0	=	12.0
1 N	–	1 × 8.0	=	8.0
				78.5 BTU/mole °F

$$C_p \text{ D.E.A.} = \frac{78.5}{105} = 0.74 \text{ BTU/Lb. °F}$$

$$C_p(\text{mixture}) = 0.5 \times 0.74 + 0.5 \times 1 = 0.37 + 0.5$$

$$C_p(\text{mixture}) = 0.87 \text{ BTU/Lb °F}$$

USE $C_p = 0.9$ BTU/LB. °F

Viscosity D.E.A.

$$\mu(20\% \text{ D.E.A.}) = 1.75\, \mu_{H_2O}$$

$$\ln \phi = w_{DEA} \ln \phi_{DEA} + w_{H_2O} \ln \phi_{H_2O}$$

$$\frac{1}{1.75\, \mu_{H_2O}} = 0.572\, \phi_{H_2O}$$

$$\ln 0.572 + \ln \phi_{H_2O} = 0.2 \ln \phi_{DEA} + 0.8 \ln \phi_{H_2O}$$

$$0.2 \ln \phi_{DEA} = \ln 0.572 + 0.2 \ln \phi_{H_2O}$$

$$\ln \phi_{DEA} = \frac{-0.559}{0.2} + \ln \phi_{H_2O} = \ln \phi_{H_2O} - 2.79$$

$$\phi_{DEA} = \phi_{H_2O}\, e^{-2.79} = \frac{\phi_{H_2O}}{16.3}$$

for 50% D.E.A. solution

$$\ln \phi = 0.5 \ln \frac{\phi_{H_2O}}{16.3} + 0.5 \ln \phi_{H_2O}$$

$$\ln \phi = 0.5(-2.79) + 0.5 \ln \phi_{H_2O} + 0.5 \ln \phi_{H_2O}$$

$$\ln \phi = -1.395 + \ln \phi_{H_2O}$$

$$\phi = \frac{\phi_{H_2O}}{e^{1.395}} = \frac{\phi_{H_2O}}{4.04}$$

$$\mu = \frac{4.04}{\phi_{H_2O}} = \underline{\underline{4.04\,\mu_{H_2O}}} \qquad 4.04 \times 2.42 = \underline{\underline{9.78}}$$

Thermal Conductivity

$k_{DEA} \cong 0.08$

$k_{H_2O} \cong 0.38$

$k(50\%\ D.E.A.) = 0.5 \times 0.08 + 0.5 \times 0.38 = 0.04 + 0.19 = \underline{\underline{0.23}}$

Stripping Tower Feed Preheater

Heat with 30 psig steam $T = 274°F$

Use: 1/2" OD X 18 BWG Tubes
$Re > 10{,}000$
steam on shell side

Estimate Tubes per Pass (h)

$$\frac{DG}{\mu} = 10{,}000 = \frac{0.4\,G}{12 \times 9.78 \times 0.5}$$

$$G_{minimum} = \frac{12 \times 9.78 \times 0.5}{0.4} \times 10^4 = 147 \times 10^4$$

$$h_{max.} = \frac{43,610 \times 144}{147 \times 10^4 \times 0.1269} = 10.5$$

<u>h = 10 Tubes per pass</u>

$$G = \frac{13,610 \times 144}{10 \times 0.1269} = 1.544 \times 10^6$$

$$T_{av} = \frac{95 + 205}{2} = 150$$

$$\mu_{150} = 9.78 \times 0.432 = 4.225$$

$$Re = \frac{0.402 \times 1.554 \times 10^6}{12 \times 4.225} = 1.224 \times 10^4$$

<u>Estimate average film temperature</u> (T_s)

$$\frac{h_i D}{k} \cong 0.027 \left(\frac{D.G.}{\mu}\right)^{0.8} \left(\frac{C_p \mu}{k}\right)^{1/3}$$

$$\frac{C_p \mu}{k} = \frac{0.9 \times 4.225}{0.23} = 16.5$$

$$h_i \cong \frac{0.027 \times 0.23 \times 12}{0.402} (12,250)^{0.8} (16.5)^{1/3}$$

$$h_i \cong 0.1855 (1860)(2.545) = 878 \text{ BTU/hr °F ft}^2 \text{ inside}$$

h_o (condensing steam) = 2500 Btu/hr °F ft² inside

$$\Delta_m = \frac{(274-95)-(274-205)}{\ln \frac{274-95}{274-205}} = \frac{179-69}{\ln \frac{179}{69}}$$

$$\Delta_m = \frac{110}{\ln 2.6} = \frac{110}{0.956} = 115\,^\circ F$$

$T_s \quad 274 - \Delta_o$

$$\Delta_o = \frac{115 \times 878}{2500 + 878} = 30\,^\circ F$$

$T_s = 254\,^\circ F$

$\mu_s = 9.78 \times 0.230 = 2.250$

Tube Side Coefficient (h_1)

$$\frac{h_1 D}{K} = 0.027 \left(\frac{\mu}{\mu_s}\right)^{0.14} \left(\frac{DG}{\mu}\right)^{0.8} \left(\frac{C_p \mu}{k}\right)^{1/3}$$

$$\left(\frac{\mu}{\mu_s}\right)^{0.14} = \left(\frac{4.225}{2.250}\right)^{0.14} = (1.877)^{0.14} = 1.092$$

$$h = 878 \times 1.092 = 958$$

Overall Coefficient (U)

$$U = \frac{h_1 h_o}{h_1 + h_o} = \frac{958 \times 2500}{958 \times 2500} = \frac{958 \times 2500}{3458} = 693 \text{ Btu/hr } ^\circ F \text{ ft}^2$$

Area Required (A)

Heat load $Q = (205-95)\ 13{,}610 \times 0.9 = 1{,}348{,}000$ Btu/hr

$$A = \frac{1{,}348{,}000}{693 \times 115} = 16.9 \text{ ft}^2 \text{ (inside)}$$

Length of Exchanger (L) use 10% Fouling Factor

Area/foot = 0.1052 X 10 = 1.052 ft²/foot

$$L = \frac{16.9 \times 1.1}{1.052} = 17.7 \text{ ft}$$

Use two passes 9 feet long

Pressure Drop through Tubes

$$\text{Velocity (v)} = \frac{13,610 \times 144}{10 \times 65.35 \times 3600 \times 0.1269} = 6.57 \text{ ft/sec.}$$

Friction Factor (f) = 0.0075 at Re = 12,240

$$F = \frac{2 f V^2 L}{g_2 D} = \frac{2 \times 0.0075 \times (6.57)^2 \times 20 \times 12}{32.2 \times 0.402}$$

F = 12.05 ft Lb/Lb

$$\Delta P = \frac{F\rho}{144} = \frac{12.05 \times 65.35}{144} = 5.5 \text{ psi}$$

Summary of Preheater Design

Length = 9 ft.

Tubes = 20 - 1/2" x 18 BWG

Passes = 2 - 10 tubes per pass

Area (A) = 9 x 20 x 0.1052 = 18.9 ft^2 (inside)

Heat load (Q) = 1,350,000 Btu/hr.

Average coefficient (U_{av}) = 693 Btu/hr °F ft^2 (inside)

Pressure Drop (ΔP) = 5.5 psi

Operation: 30 psig steam on shell side
T = 274°F

Steam Required = $\frac{1.35 \times 10^6}{928.6}$ = 1455 Lb/hr

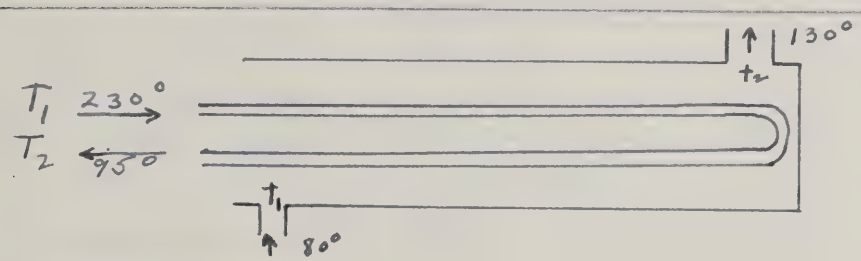
LEAN SOLUTION COOLER

Use: 2 tube passes, 1 shell pass
1/2" OD x 18 BWG Tubes
Cooling water: on shell side
in at 80°F out at 130°F

$Q = 13,610 \times (230-95) \times 0.9 = 1,653,000$ BTU/hr

Water Rate $(w) = \dfrac{1,653,000}{(130-80)} = 33,060$ Lb/hr.

$w = 3970$ gal/hr.



$$\Delta_m = \frac{(230-130)-(95-80)}{\ln \frac{230-130}{95-80}} = \frac{100-15}{\ln \frac{100}{15}} = \frac{85}{\ln 6.67}$$

$$\Delta_m = \frac{85}{1.896} = 44.8°F$$

$$Z = \frac{T_1 - T_2}{t_2 - t_1} = \frac{230-95}{130-80} = \frac{135}{50} = 2.7$$

$$X = \frac{t_2 - t_1}{T_1 - t_1} = \frac{130-80}{230-80} = \frac{50}{150} = 0.333$$

$F = 0.5$ from Fig 64-A, P146 McAdams

Thus it is obvious that the use of two passes is futile, since $\Delta = 1/2\,\Delta_m$ but $\Delta_{2\,passes} = 2A_{1\,pass}$ the heat transferred will be the same in each case.

Use: 1 tube pass, 1 shell pass

Estimation of Wall Temperature

Inside Fluid

$$T_{av} = \frac{230+95}{2} = 163^\circ$$

$$\mu_{163^\circ} = 9.78 \times 0.391 = 3.83$$

$$Re = \frac{0.402 \times 1.554 \times 10^6}{12 \times 3.83} = 1.361 \times 10^4$$

$$h_i \cong 0.1855 \left(\frac{DG}{\mu}\right)^{0.8} \left(\frac{C_p \mu}{k}\right)^{1/3}$$

$$\frac{C_p \mu}{K} = \frac{0.9 \times 3.83}{0.23} = 15.0$$

$$h_i \cong 0.1855 (13,610)^{0.8} (15)^{1/3}$$

$$h_i \cong 0.1855 (2005)(2.465) = 917$$

Shell Side Fluid

$$T_{av} = \frac{130+80}{2} = 105^\circ F$$

$$\mu_{150^\circ} = 2.42 \times 0.649 = 1.57$$

Maximum Free Area = $B \Sigma p$

B = Baffle Spacing = 8"
P = Free Length between tubes

$$B \Sigma p = 8 \times 1/4'' \times 5 = 10 \text{ in}^2$$

$$G = \frac{33,660 \times 144}{10} = 476,000 \text{ lb/hr ft}^2$$

$$Re = \frac{DG}{\mu} = \frac{0.5 \times 4.76 \times 10^5}{12 \times 1.57} = 12,600$$

$$h_o \cong \frac{0.33 \times 0.36 \times 12}{0.500} \left(\frac{DG}{\mu}\right)^{0.6} \left(\frac{C_p \mu}{K}\right)^{1/3} L_f$$

L_f = Leakage factor = 0.6

$$\frac{C_p \mu}{K} = \frac{1 \times 1.57}{0.36} = 4.36$$

$$h_o \cong 1.711 \, (12600)^{0.6} \, (4.36)^{1/3}$$

$$h_o \cong 1.711 \times 287 \times 1.634 = 802$$

$$T_{wall} = 163 - \Delta$$

$$\Delta = \frac{\Delta_m h_o}{h_o + h_i} = \frac{45 \times 802}{917 + 802} = 21°F$$

$$T_{wall} = 163 - 21 = 142°F$$

<u>Inside Coefficient h_i</u>

$$h_i = 0.1855 \left(\frac{DG}{\mu}\right)^{0.8} \left(\frac{C_p \mu}{K}\right)^{1/3} \left(\frac{\mu}{\mu_s}\right)^{0.14} = 917 \left(\frac{\mu}{\mu_s}\right)^{0.14}$$

$$\mu_{142} = 9.78 \times 0.461 = 4.51$$

$$\left(\frac{\mu}{\mu_s}\right)^{0.14} = \left(\frac{3.83}{4.51}\right)^{0.14} = (0.85)^{0.14} = 0.978$$

$$\underline{\underline{h_i = 917 \times 0.978 = 897 \text{ BTU/hr °F ft}^2 \text{ inside}}}$$

Shell Side Coefficient

$T_w = 142°F$

$T_{av} = 105°F$

$T_f = \frac{142+105}{2} = 123°F$

$$h_o = \frac{0.33\, k_f \times 12}{0.500} \left(\frac{DG}{\mu_f}\right)^{0.6} \left(\frac{c_p \mu_f}{k_f}\right)^{1/3} L_f$$

$k_f = 0.37$

$\mu_f = 2.42 \times 0.544 = 1.316$

$$\left(\frac{DG}{\mu_f}\right)^{0.6} = \left(\frac{0.5 \times 4.76 \times 10^5}{12 \times 1.316}\right)^{0.6} = (15{,}020)^{0.6} = 320$$

$$\left(\frac{c_p \mu_f}{k_f}\right)^{1/3} = \left(\frac{1 \times 1.316}{0.37}\right)^{1/3} = (3.55)^{1/3} = 1.525$$

$h_o = 1.758 \times 320 \times 1.525 = 858$ Btu/hr °F ft² (outside)

$h_o = \frac{858 \times 0.5}{0.402} = 1066$ Btu/hr °F ft² Inside

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$$U = \frac{1066 \times 897}{1066 + 897} = \frac{1066 \times 897}{1963} = 487 \text{ Btu/hr °F ft}^2 \text{ (inside)}$$

Area Required

$$A = \frac{1{,}653 \times 10^6}{487 \times 44.8} = 75.8 \text{ ft}^2$$

Length of Exchanger

$$L = \frac{75.8}{10 \times 0.1052} = 72.2 \text{ ft.}$$

Use 5-15 foot exchangers In Series

Summary of Cooler Design

Length = 15 feet

Number = 5 Units

Tubes = 10 per exchanger 1/2" x 18 BWG

Area = $5 \times 10 \times 15 \times 0.1052 = 78.9\ ft^2$

Baffle Spacing = 10"

Heat Load = 1.653×10^6 Btu/hr

Average Coefficient (U_{av}) = 487 Btu/hr °F ft^2 (inside)

Cooling H_2O Rate = 33,060 Lb/hr = 3970 gal/hr.

Pressure Drop (ΔP) = 21 psi

CONDENSER

Heat Load

$Q_{H_2S} = 14.3 \times 7 \times (205-123) = 8100$ Btu/hr

$Q_{H_2O} = 2270 \times (1148-91) = 2,410,000$ Btu/hr

$Q_{Total} = 2,420,000$ Btu/hr

Cooling Water

$T_{inlet} = 80°F$
$T_{outlet} = 120°F$

$$W = \frac{2.42 \times 10^6}{40} = 6.05 \times 10^4 \text{ Lb/hr} = 7260 \text{ gph}$$

Water rate is made large by use of low outlet temperature in order to reduce deposition of silt and obtain large Δ and large h_w.

Inlet Conditions

Water Vapor pressure, $P_{vi} = 0.869$ Atm.
Inert Gas pressure, $P_{gi} = 0.368$ Atm.

Exit Conditions

$P_{vo} = 0.124$ Atm.
$P_{go} = 1.113$ Atm.

Steam condensed = Reflux = 126 Moles/hr

Use -19 Tubes 1/2" O.D. x 18 BWG

Heat Transfer Coefficients

Condensate film $h_c = 2000$ (Assumed safe value)

Metal tube $h_p = \frac{65 \times 12}{0.049} = 15{,}900$

Dirt and Scale $h_d = 700$ (Assumed)

Water film coefficient $\frac{h_w}{L_f} = \frac{370\,(1 + 0.0067 t_f)(V'_{max})^{0.6}}{(D_o')^{0.4}}$

t_f = film temp. °F = 120°F
V'_{max} = velocity feet/sec
D_o' = diameter inches
L_f = leakage factor = 0.6

Baffle Spacing = 8"

Free Area $= \frac{6}{4} \times \frac{8}{144} = \frac{1}{12}$ ft²

$V' = \frac{6.05 \times 10^4 \times 12}{62.4 \times 3600} = 3.22$ ft/sec.

$h_w = \frac{0.6 \times 370\,(1 + 0.8)(3.22)^{0.6}}{(0.5)^{0.4}}$

$h_w = \frac{222\,(1.8)(2.016)}{0.758} = 1060$

$R = \frac{1}{2000} + \frac{0.402}{15900 \times 0.451} + \frac{0.402}{700 \times 0.500} + \frac{0.402}{1060 \times 0.500}$

$R = 0.0005 + 0.0000457 + 0.00114857 + 0.000758$

$R = 0.002453$

$h_o = 408$ Btu/hr °F ft² (inside)

Average $U = 300$ Btu/hr °F ft² inside

by analogy with example given in "Design of Cooler Condensers for Mixtures of Vapors with Non-Condensing Gases"
Colburn, A.P. & Hougen, O.A.
I. & E.C., Vol. 26, Page 1178, November 1934

300 Btu/hr °F ft² inside is conservative since fraction of Inert gas is $\frac{14.3}{140.3} = 0.102$ Moles inert/mole gas

Compared with 0.165 Moles inert/mole gas in illustrative example.

Similarity to illustrative example is shown by following data.

	example	present problem
h_o outside area	310	$\frac{408 \times .402}{0.500} = 328$
h_o inside area	$\frac{310 \times .75}{0.62} = 375$	408
Moles condensed	498	126
No. of Tubes	83	19
$\frac{\text{Moles condensed}}{\text{Tube}}$	$\frac{498}{83} = 6$	$\frac{126}{19} = 6.6$

Area Required

$$\Delta_m = \frac{(205-120)-(123-80)}{\ln \frac{205-120}{123-80}} = \frac{85-43}{\ln \frac{85}{43}}$$

$$\Delta_m = \frac{42}{\ln 1.977} = \frac{42}{0.682} = 61.6\,^\circ F$$

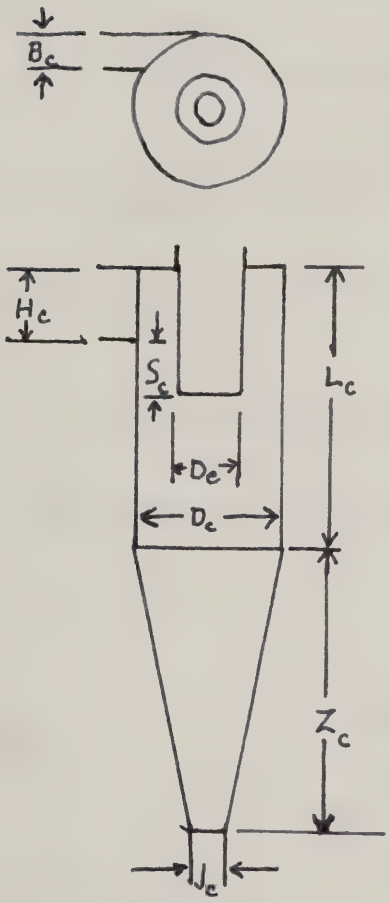
$$A = \frac{2.42 \times 10^6}{61.6 \times 300} = 131\ ft^2 \text{ (inside)}$$

$$\text{Length} = \frac{131}{19 \times 0.1052} = \underline{\underline{72\ ft.}}$$

CYCLONE SEPARATOR

Minimum Particle Diameter $= 20\mu = 20 \times 10^{-4}$ cm

$$= \frac{20 \times 10^{-4}}{30.48} = 6.56 \times 10^{-5} \text{ ft.}$$



$B_c = D_c/4$

$D_e = D_c/2$

$H_c = D_c/2$

$L_c = 2D_c$

$S_c = D_c/8$

$Z_c = 2D_c$

$J_c = D_c/4$

Inlet Area $= B_c H_c = D_c^2/8$

$$N_t = \frac{L_c}{H_c} = \frac{2D_c}{D_c/2} = 4$$

Inlet Gas Velocity

$$\text{Flow Rate} = 2.567 \times 10^6 \times \frac{14.7 \times 555}{315 \times 492} = 1.35 \times 10^5 \ \frac{ft^3}{hr.}$$

$$V_c = \frac{1.351 \times 10^5 \times 8}{D_c^2} = \frac{1.081 \times 10^6}{D_c^2} \ ft/hr$$

Viscosity of Gas

Viscosity at present temperature and pressure is not affected by pressure ∴

μ_{CH_4} (95°F) = 0.0112 × 2.42 = 0.0271 Lb/hr ft

$\mu_{C_2H_6}$ (95°F) = 0.0094 × 2.42 = 0.0227 Lb/hr ft

$$\phi_{CH_4} = \frac{1}{0.0271} = 36.9$$

$$\phi_{C_2H_6} = \frac{1}{0.0227} = 44.1$$

$$\ln \phi = 0.8 \ln 36.9 + 0.2 \ln 44.1$$

$$\ln \phi = 0.8 \times 3.61 + 0.2 \times 3.79 = 2.885 + 0.757 = 3.642$$

$$\phi = 38.2$$

$$\mu = \frac{1}{38.2} = 0.0262 \text{ Lb/hr ft.}$$

Density of Gas

$$\rho = \frac{PM}{RT} = \frac{315 \times 18.44}{14.7 \times 0.729 \times 555} = 0.977 \text{ Lb/ft}^3$$

Cyclone Diameter

$$D_{p\,min} = \sqrt{\frac{9\mu B_c}{\pi N_T V_c (\rho - \rho_g)}} = \sqrt{\frac{9\mu D_c \times D_c^2}{4\pi \times 4 \times 1.081 \times 10^6 (\rho - \rho_g)}}$$

$$D_{p\,min}^2 = \frac{9 \times 0.0262 \text{ Lb/hr ft } D_c^3}{16\pi \times 1.081 \times 10^6 \text{ ft}^3/\text{hr } (65.35 - 0.98) \text{ Lb/ft}^3} = (6.56 \times 10^{-5})^2 \text{ ft}^2$$

$$D_c^3 = \frac{43.03 \times 10^{-10} \times 16\pi \times 1.081 \times 10^6 (64.37)}{9 \times 0.0262} = 63.83 \text{ ft}^3$$

$$D_c^3 = 63.83 \text{ ft}^3$$

$$D_c = \sqrt[3]{64 \text{ ft}} = 4 \text{ ft.} = 48 \text{ inches}$$

<u>Make $D_c = 48$ inches</u>

<u>Pressure Drop</u>

$D_c = 4$ ft
$B_c = 1$ ft
$H_c = 2$ ft
$D_e = 2$ ft
$L_c = 8$ ft
$Z_c = 8$ ft
$J_c = 1$ ft

$$F_{cv} = \frac{16 B_c H_c}{D_e^2}$$

$$F_{cv} = \frac{16 \times 1 \times 2}{4} = 8$$

Inlet Velocity $V_c = \dfrac{1.351 \times 10^5}{1 \times 2 \times 3600} = 18.8$ ft/sec

Inlet velocity head $= \dfrac{V_c^2 \rho}{2g \times 144}$ Lb/in^2

$$\Delta P = \frac{8 V_c^2 \rho}{288 g} = \frac{V_c^2 \rho}{36 g} = \frac{(18.8)^2 \times 0.977}{36 \times 32.2} = 0.299$$

$\Delta P \approx 0.3$ psi

Thickness of wall $= \dfrac{PD}{2SE} = \dfrac{315 \times 48}{2 \times 14{,}000} = 0.54''$

make $t = 0.625''$

Summary of Cyclone Design

Diameter	4 feet
Exit Diameter	2 feet
Overall Height	16 feet
Inlet Area	2 feet^2
Wall Thickness	0.625"
Inlet Velocity	18.8 ft/sec.
Pressure Drop	0.3 psi

PUMP DESIGNS

Main Pump

$\Delta P = 350$ psi

$$\text{Work} = \frac{350 \times 144}{65.35} = 771 \text{ ft Lb/Lb}$$

$$\text{Power} = \frac{771 \times 13,610}{1.98 \times 10^6} = 5.3 \text{ Horsepower}$$

$E = 50\%$

$$\frac{5.3}{0.5} = 10.6 \text{ Horsepower input}$$

Motor Efficiency = 90%

$$\frac{10.6}{0.9} = 11.8 \text{ Horsepower} = \underline{8.9 \text{ KW}}$$

Reflux Pump

$\Delta P = 50$ psi

$$\text{Work} = \frac{50 \times 144}{62.4} = 115 \text{ ft Lb/Lb}$$

$$\text{Power} = \frac{115 \times 2270}{1.98 \times 10^6} = 0.132 \text{ HP}$$

Efficiency = 10%

$$\text{Power input} = \frac{0.132}{0.10} = 1.3 \text{ HP}$$

$$\frac{1.3}{0.75} = 1.74 \text{ HP} = \underline{\underline{1.31 \text{ KW}}}$$

D.E.A. SOLUTION SURGE TANK

Total Liquid System

1/ Absorption Tower

Height of liquid on plates = 6"

Height of liquid in bottom = 20"

Total liquid height = $\frac{27 \times 6 \times 20}{12}$ = 15.2 ft.

D = 7 ft.

$V_1 = \frac{49\pi \times 15.2}{4} = 585 \text{ ft}^3$

2/ Stripping Tower

Height of liquid on plates = 6"

Height of liquid in bottom = 30"

Total Liquid Height = $\frac{60 \times 6 + 30}{12}$ = 32.5 ft

D = 3 ft

Liquid in Reboiler = 10% Liquid in column

$V_2 = 1.1 \times \frac{9\pi}{4} \times 32.5 = 353 \text{ ft}^3$

3/ Heat Exchangers

Total Tube length = 20×10 + 75×10 = 950 ft.

$V_3 = \frac{950 \times 0.1052}{144} = 0.7 \text{ ft}^3$ Negligible

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4/ Surge Tank Running Volume

V_4 = 25% of hourly circulation

$$V_4 = 0.25 \times \frac{13{,}610 \text{ Lb/hr}}{65.4} = 208 \text{ ft}^3$$

Assume: Liquid in Lines = 10% Total Liquid

$$\text{Total Liquid} = \frac{585 + 353 + 208}{0.9} = \frac{1146}{0.9} = 1275 \text{ ft}^3$$

Volume of Surge Tank use 1.3 safety factor

$$V_s = 1275 \times 1.3 = 1660 \text{ ft}^3$$

Use: 15 ft Diameter

$$h = \frac{1660 \times 4}{225\pi} = \underline{9.4 \text{ ft.}}$$

FEED GAS SURGE TANK

L = 15 feet

D = 10 feet

V = 78.5 × 15 = 1178 ft^3

COSTS OF EQUIPMENT

D.E.A. Surge Tank

$P = h\rho = 10 \times 65.35 = 653.5\ Lb/ft^2 = 4.6\ Lb/in^2$

$S\ (cast\ iron) = 23000\ Lb/in^2$

$S\ (allowable) = \frac{23000}{5} = 4600\ Lb/in^2$

$t = \frac{4.6 \times 15 \times 12}{2 \times 4600} = 0.090$ inches

Make it 0.125 inches

$W = 437 \times \left(10 \times 15\pi + \frac{2 \times 15^2 \pi}{4}\right) \frac{0.125}{12}$

$W = \frac{54.6\pi}{12}(150 + 112.5) = 3750\ Lb.$

Cost $= 0.2 \times 3750 =$ \$750
Erection cost $= 1600 \times .25 =$ 188
Total Cost $=$ \$938

Feed Gas Surge Tank

$P = 315\ psi$, $S\ (steel) = 70{,}000\ Lb/in^2$
$S\ (allowable) = \frac{70{,}000}{5} = 14{,}000\ Lb/in^2$

$t = \frac{315 \times 10 \times 12}{2 \times 14{,}000} = 1.35$ inches

$W = 490 \times \left(15 \times 10\pi + \frac{2 \times 10^2 \pi}{4}\right) \frac{1.35}{12}$

$W = \frac{490 \times 1.35\pi}{12}(150 + 50) = 34{,}600\ Lb.$

Unit Cost $= 0.2 \times 34{,}600 =$ \$6920
Erection Cost $= 6920 \times 0.25 =$ 1730
Total Cost $=$ \$8650

Absorption tower

For 7 foot diameter plates (low pressure)
Cost of bubble cap unit = \$650
Assume shell thickness = 1/4 inch

$$\text{Shell wt.} = \frac{490 \times 2 \times 7\pi \times 0.25}{12} = 450 \text{ Lb.}$$

Cost of low pressure shell = 0.20 × 450 = \$90
Cost of plate without shell = 650 − 90 = \$560
for high pressure 315 psi

$$t = \frac{315 \times 7 \times 12}{2 \times 14{,}000} = 0.945 \text{ inches}$$

$$W = \frac{490 \times 2 \times 7\pi \times 0.95}{12} = 1710 \text{ Lb.}$$

Shell cost = 0.2 × 1710 = \$342

Cost of bubble cap unit = 342 + 560 = \$902
Erection cost = 0.4 × 902 = 361
Total cost per bubble cap = \$1263

Cost of Tower = 27 × 1263 = \$34,100

Stripping Tower

Cost of bubble cap units (low pressure) = \$240
Erection cost = 0.4 × 240 = 96
Cost per plate \$336
Total cost = 60 × 336 = \$20,150

Reboiler

Cost of shell = 100 × 5.60 = \$560
Cost of tubes = 100 × 8.00 × 1.25 × 1.2 = 1200
Total \$1760

Installation cost = 0.4 × 1760 = 704

Total cost of Reboiler = \$2464

Stripping Tower Feed Heater

Cost of shell = 21 × 5.60 =	$118
Cost of tubes = 21 × 8.00 × 1.25 =	210
Total	$328
Installation = 0.15 × 328	49
Total cost of heater	$377

Lean Solution Cooler

Shell cost = 80 × 5.6 × 1.2 =	$537
Tube cost = 80 × 8 × 1.25 × 1.2 =	960
Installation cost = 0.15 (537 + 960) =	225
Total =	$1722

Stripping Tower Condenser

Shell cost = 131 × 5.6 =	$734
Tube cost = 131 × 8 × 1.25 =	1310
Installation cost = 0.25 (1310 + 734) =	411
Total =	$2455

Main Pump

Pump Cost = 11 × 45 =	$495
Motor Cost = 11 × 6 =	66
Installation = 0.12 × (495 + 66) =	67
Total per pump =	628
Stand by	628
Total	$1256

Reflux Pump

Pump cost	1.3 x 150 =	$195
Motor cost	=	40
Installation = 0.15 x (195 + 40)		35
Total cost per pump		$ 270
Stand by unit		270
Total		$ 540

Cyclone Separator

Neglect exit Duct, calculate volume as that of a cylinder 16' long x 4 ft Dia.

$$V = 16 \times 4\pi \times \frac{0.625}{12} = 10.47 \text{ ft}^3$$

Weight = 490 x 10.47 = 5130 Lb.

Cost = 0.4 x 5130 = $2052

Installation = 2052

Total = $4104

BASIC EQUIPMENT COSTS

ITEM	INVESTMENT
D.E.A. Surge Tank	$ 940
Feed Gas Surge Tank	8,650
Absorption Tower	34,100
Stripping Tower	20,150
Reboiler	2,465
D.E.A. Heater	371
D.E.A. Cooler	1,722
Condenser	2,455
Main Pump	1,260
Reflux Pump	540
Cyclone Separator	4,104
Total Basic Equipment Cost	$ 76,763

Auxiliary Equipment Costs

ITEM	% Basic Equipment Cost
Foundation Material	1
Building Material	3
Structural Steel	2
Piping	18
Electrical Wiring	1/2
Instrumentation	4 1/2
Insulation, Fireproofing & Paint	1
TOTAL	30% Basic Equipment Cost

Auxiliary Equipment Cost = 0.3 × 76,763 = $23,030

TOTAL Constructed Cost = 76,763 + 23,030 = $99,793

YEARLY COSTS

Taxes + Insurance	— 0.015 × 99,793 =	$1,497
Interest	— 0.045 × 99,793 =	4,491
Overhead	— 0.060 × 99,793 =	5,988
		$11,976

Operating Costs

Heating steam, is taken from turbines therefore cost will be calculated for turbines and need not be considered here.

Cooling Water

Cooler —	3970 gph
Condenser —	7260 gph
Total	11,230 gph

$$\text{Cost/hr} = \frac{11{,}230}{1000} \times 0.05 = \$0.567/\text{hr.}$$

Electric Power

Reflux Pump	1.3 kw
Main Pump	8.9 kw
Total	10.2 kw

$10.2 \times 3413 = 34{,}820$ Btu/hr.

ΔH_s (Steam 65 psia to 2 psia) = 375 Btu/Lb.

$\Delta H_{actual} = 0.75 \times 375 = 281$ Btu/Lb.

Generator Efficiency = 0.95

$$\text{Steam Required } \frac{34{,}820}{0.95 \times 281} = 131 \text{ Lb/hr.}$$

$$\text{Power Cost} = \frac{131 \times .45}{1000} = \$0.059/\text{hr.}$$

Operators = $2.000/hr.

SUMMARY OF HOURLY COSTS

		$/hr
Water		0.567
Power		0.057
Operators		2.000
D.E.A.		0.250
Fixed Charges	18,376 / 8760	2.100

Main Compressors

The main oxygen and natural gas compressors will be multi-stage, turbine driven, centrifugal compressors of an advanced design. The oxygen compressor will have four stages, operating at a compression ratio of 2.2, with intercooling but no aftercooling. The natural gas compressor will have one stage, operating at a compression ratio of 2.2, with a Venturi scrubber as an aftercooler in order that the natural gas be delivered to the H_2S removal unit saturated with water vapor at 95 °F.

Horsepower:	
Oxygen compressor	13,650
Natural gas compressor	4,500
Turbine	18,150
Pump for Venturi scrubber	210
Steam required	~~767~~ 167,000 #/hr.
Cooling water required	716,000 gal/hr.
Total Cost	$2,300,000

Main Oxygen compressor:

Thruput = 83 tons/hr
= 166,000 #/hr.
= 1,860,000 ft^3/hr.

critical constants for O_2

$t_c = -119°C = 277°R$
$p_c = 50$ atm.

ranges of reduced temperatures and pressure

$$T_r = \frac{560}{277} = 2.0 \quad \text{to} \quad \frac{760}{277} = 2.7$$

$$p_r = \frac{1}{50} = 0.02 \quad \text{to} \quad \frac{22}{50} = 0.44$$

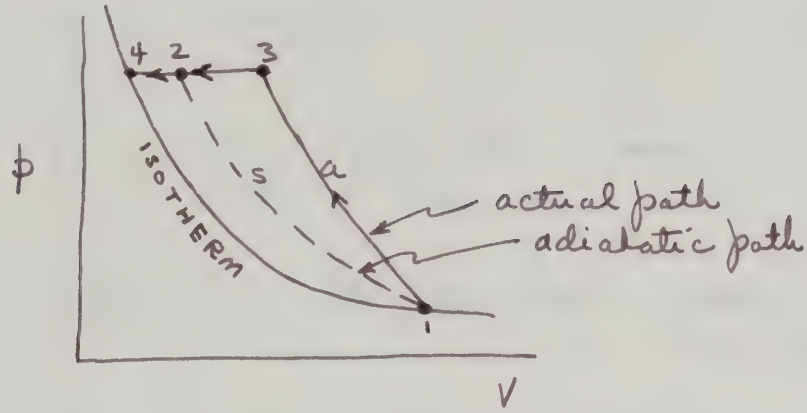
$$C = \frac{pV}{RT} = 1 \quad \text{over entire range}$$

therefore work may be calculated as adiabatic compression of an ideal gas. with $k = 1.4$

A 4-stage centrifugal compressor with intercoolers, powered by a direct drive turbine will be used.

$$\frac{p_2}{p_1} = 2.2 \quad \text{for each stage}$$

Each stage is assumed to follow a path as shown.



Equations for calculating work and heat effects per mol for each stage

$$W_s = \frac{RT_1}{k-1}\left[1-\left(\frac{p_2}{p_1}\right)^{\frac{k-1}{k}}\right]$$

$$= \frac{1.99\ T_1}{0.4}\left[1-(2.2)^{0.286}\right]$$

$$= 5\ T_1\left(1-1.253\right)$$

$$= -5\left(0.253\right)T_1 = 1.265\ T_1 \text{ Btu}$$

$$W_a = \frac{1.265\ T_1}{0.60} = 2.11\ T_1 \text{ Btu}$$

assuming a 60% efficiency

$$T_2 = T_1 \left(\frac{p_2}{p_1}\right)^{0.286}$$

$$= 1.253\, T_1$$

$$T_3 = T_2 + \frac{W_a - W_s}{7.15} = \cancel{1.25}$$

where 7.15 Btu/mol × °F = molar specific heat of O_2

$$T_3 = 1.253\, T_1 + \frac{(2.11 - 1.265)\, T_1}{7.15}$$

$$= \left(1.253 + \frac{0.845}{7.15}\right) T_1$$

$$= 1.371\, T_1$$

$$W_{3-4} = \cancel{p_2}(V_4 - V_3)$$

$$= R(T_4 - T_3)$$

$$= -1.99(T_3 - T_4)$$

$$Q_{3-4} = 7.15\,(T_4 - T_3)$$

$$= -7.15\,(T_3 - T_4)$$

STAGE	p_1 psia	p_2 psia	T_1 °R	T_4 °R	W_a Btu	W_{3-4} Btu
1	14.7	32.3	560	767	1180	313
2	31.3	68.8	610	836	1286	450
3	66.8	147	610	836	1286	450
4	145	319	610	836	1286	450

STAGE	W_{TOTAL} Btu	Q Btu
1	1493	1123
2	1736	1615
3	1736	1615
4	1736	
	6701 Btu/mol	4353 Btu/mol.

Total work done per mol = 6701 Btu/mol.

Total heat removed in intercoolers = 4353 Btu/mol.

$$\text{mols/hr throughput} = \frac{166,000}{32} = 5190$$

$$\text{HP of drive turbine} = \frac{(6701)(5190)}{(2545)} = 13,650$$

$$\text{Intercooler heat load} = (4353)(5190) = 22,600,000 \text{ Btu/hr.}$$

$$\text{Cooling water required} = \frac{22,600,000}{(200-80)} = 188,000 \text{ \#/hr.}$$

$$= 22,600 \text{ gal/hr.}$$

Discharge pressure = 319 psia.

Discharge temperature = 376 °F

Natural Gas Compressor

Natural Gas is supplied at 150 psia, and a single stage compressor operating at a $\frac{p_2}{p_1} = 2.2$ will be used. It will be driven by the same turbine as the oxygen compressor.

Pseudo-critical constants for natural gas.

$$T_c = (0.8)(191.1) + (0.2)(305.2) =$$
$$= 152.8 + 61.0$$
$$= 212.2\,°K = 383\,°R$$

$$p_c = (0.8)(45.8) + (0.2)(48.8)$$
$$= 36.6 + 9.8$$
$$= 46.4 \text{ atm.}$$

$$T_r = \frac{560}{383} = 1.46 \text{ to } \frac{760}{383} = 1.98$$

$$p_r = \frac{10}{46.4} = 0.216 \text{ to } \frac{22}{46.4} = 0.474$$

$$C = 0.98 \text{ to } 1.00$$

Natural gas may also be treated as a perfect gas.

$$k = (0.8)(1.31) + (0.2)(1.19)$$
$$= 1.048 + 0.238$$
$$= 1.286 = 1.29$$

$$W_s = \frac{1.99\,T_1}{1.29 - 1}\left[1 - (2.2)^{\frac{1.29-1}{1.29}}\right]$$

$$= (6.86)(560)\left[1 - (2.2)^{0.225}\right]$$

$$= 3840\,(1. - 1.194)$$

$$= 745 \text{ Btu/mol}$$

$$W_a = \frac{745}{0.60} = 1240 \text{ Btu/mol}$$

$$T_2 = 1.194\,(560) = 679\ ^\circ R$$

$$T_3 = 679 + \frac{(1240 - 745)}{9.5}$$

where 9.5 Btu/mol × °F = molar specific heat of natural gas.

$$T_3 = 679 + 52 = 731\ ^\circ R$$

$W_{3-4} = 1.99\,(731-555)$ cooled in scrubber to 95°F. saturated

$= 350$ Btu/mol.

Total work $= 1240 + 350 = 1590$ Btu/mol

Heat removed in scrubber $= (731-555)(9.5)$
$= 1673$ Btu/mol.

mols/hr. thruput $= \dfrac{62,000,000}{(359)(24)} = 7200$ mols/hr.

H.P. of drive turbine $= \dfrac{(7200)(1590)}{2545} = 4500$

Scrubber heat load $= (7200)(1673) = 12,050,000$ Btu/hr.

Scrubber acts as an aftercooler.

Complete drive turbine for natural gas and oxygen compressors.

H.P. $= 13650 + 4500 = 18,150$

Power ~~Power~~ in Btu/hr $= (18,150)(2545)$

$= 46,200,000$ Btu/hr.

Turbine operates between 650 psia and 2 psia

$$\Delta H_s = (1204 - 829) = 375 \text{ Btu/\# steam}$$

$$\Delta H_a = (0.75)(375) = 281 \text{ Btu/\# steam}$$

steam required to drive compressor turbine

$$\frac{46,200,000}{281} = 165,000 \text{ \#/hr.}$$

Cost of turbine and ~~pumps~~ compressors is estimated at \$120 per H.P

$$18,150\,(120) = \$2,180,000$$
$$\cong \$2,200,000$$

Water for condensers $= (30)(167,000) = 5,010,000$ #/hr.
$= 600,000$ gal/hr.

<u>After cooler (scrubber type) for natural gas compressor.</u>

p_{H_2O} at 95°F $= 0.8153$ psia

mol fraction H_2O in gas leaving scrubber

$$\frac{0.815}{330} = 0.00247 \text{ mols } H_2O/\text{mol gas.}$$

H_2O evaporated $= 0.00247\,(7200) = 17.8$ mols/hr.
$= 321$ #/hr.

ΔH for H_2O evaporated = 1103 - 63 = 1040 Btu/#

1040 (321) = 334,000 Btu/hr.

Heat removed by cooling water = 12,050,000 - 334,000
= 11,716,000 Btu/hr.

Cooling H_2O required = $\frac{11,716,000}{95-80}$ = 780,000 #/hr.
= 93,500 gal/hr.

assume a throat velocity of 250 ft/sec for the Venturi scrubber.

$\frac{7200}{3600}$ = 2 mols/sec thru put

volume thruput = $\frac{(2)(0.729)(555)}{22.4}$ = 36 ft^3/sec.

area of throat = $\frac{36}{250}$ = 0.144 ft^2 = 20.7 in^2

diameter of throat = 5.12 in.

Water pump for the scrubber.

ΔP = 340 - 40 = 300 psia

work = $\frac{300(144)}{62.4}$ = 692 ft#/#

power = $\frac{(692)(780000)}{1,980,000}$ = 273 H.P

assume an efficiency of 60%

power input $= \frac{273}{0.60} = 455$ H.P.

a Pelton wheel and turbine drive will be used.

assume Pelton wheel has 90% efficiency

Pelton wheel H.P. $= 0.90(273) = 245$ H.P.

Required turbine H.P $= 455 - 245 = 210$ H.P.

Power req'd in Btu/hr $= 210(2545) = 534{,}000$ Btu/hr.

assume driven by same type turbine as above

steam req'd $= \frac{534{,}000}{281} = 1900$ #/hr.

Costs:

Scrubber water pump	$= (455)\,35$	=	\$15,900
Pelton wheel	$(245)\,25$	=	6,120
Turbine drive	$(210)\,15$	=	3,150
			\$25,170
Installation	$(0.12)(25{,}170)$	=	3,020
			\$28,190
Stand by unit			28,190
Total			\$56,380

Total cost of compressors, turbine and auxilliary equipment = \$2,300,000

Natural Gas Preheater Design

Composition

	%
CH_4	79.8
C_2H_6	19.5
N_2	.5

MW [illegible]

Flow

62,000,000	$\frac{\text{std ft}^3}{\text{day}}$	2,580,000	$\frac{\text{std ft}^3}{\text{hr}}$
173,000	$\frac{\text{moles}}{\text{day}}$	7,200	$\frac{\text{moles}}{\text{hr}}$
3,240,000	$\frac{\text{lb}}{\text{day}}$	135,000	$\frac{\text{lb}}{\text{hr}}$

Conditions

Temp in	60°F	520°R
Temp out	1150°F	1610°R
Pressure	300 psia	

Preliminary considerations

The flow is to be distributed between 2 furnaces with a third as a replacement. The [illegible] temperature of the furnace tubes is not to exceed 1250°F which is the maximum allowed by the Power Boiler Code, ASME 1947, paragraph 23, which requires the use of [illegible] steel tubes. The mixture of preheated natural gas and preheated [illegible] must have a temperature of 1000°F. It also seems desirable to [illegible] the [illegible] temperature below 500°F to minimize the corrosion of the tubes in [illegible] section.

[illegible]

C_p for [illegible] $= 7.5 + .005\,T \quad \frac{BTU}{mol}$

$Q = 7200 \frac{mol}{hr} \int_{520}^{160} 7.5 + .005\,T \, dT \quad \frac{BTU}{mol} = 103,[illegible] \frac{BTU}{hr}$

Q for the [illegible] $= 518,[illegible] \frac{BTU}{hr}$

$A_c = 20 \times 16 \times 4 + 16^2 = 1280 + 256 = 1530 \text{ ft}^2$

$A_r = 16^2 = A_{\text{[illegible]}} = [illegible] \text{ ft}^2$

$F_{\alpha\beta} = 1.0$

$\epsilon_\beta = \frac{\alpha_\beta \left(\frac{T_1}{T_2}\right)^4}{1 - \left(\frac{T_1}{T_2}\right)^4} =$

$\frac{(11 - .198 \times 1.08 - .05) - \left(\frac{T_1}{T_2}\right)^4 (12 + .05 \times 1.08 - .28)}{1 - \left(\frac{T_1}{T_2}\right)^4}$

$= .27$

$\epsilon_\beta = .37$ [illegible]

$$\bar{F}_{c1} = e_f\left(1 + \frac{\frac{\epsilon_f}{\epsilon_c}}{1 + \frac{\epsilon_c}{1-\epsilon_f}\frac{1}{\epsilon_c}}\right) = .33\left(1 + \frac{\frac{.33}{.63}}{1 + \frac{.33}{.63}}\right) = .385$$

$$\bar{F}_{c1} = \frac{1}{\frac{1}{\epsilon_c} + \frac{1}{\epsilon_c} - 1} = \frac{1}{\frac{1}{.385} + \frac{1}{.3} - 1} = .385$$

n assumed to be 2 for [illegible]

$$Q = \frac{1}{\frac{1}{\epsilon} + \frac{1}{\epsilon} - \frac{1}{0.9}} = .47$$ for Transmission to outside

Note equation (skin temperature at out [illegible])

$$\frac{Q}{A_c} = .173 \quad \bar{F}_{c1}\left[\left(\frac{T_s}{100}\right)^4 - \left(\frac{T_s}{100}\right)^4\right] = u \quad \frac{\pi\, D\, \text{in}\, L}{D\, \text{out}\, L}\,(T_s - T_{out})$$

at 1200°F ... S (allowable [illegible] for [illegible] [illegible] 3600 psi

$$t = \frac{D\,P}{2S} + .065 = \frac{2.875 \quad 800}{2 \quad 3600} + .065 = .185$$

Std $2\frac{1}{2}$ in. pipe will be used

D out = 2.875"

D in = 2.469

T_s = 1200°F

$$T_F = 100\sqrt[4]{\frac{u_1 + 37.4}{.00492}}$$

Assume [illegible] at the outlet

N_2 = 40

$$n = \frac{2\,588000\,\frac{580}{\text{[illegible]}}\,\frac{1}{20}\,\frac{\text{ft}^3}{\text{sec}}}{2\,.0332\,\frac{\text{ft}^2}{\text{pipe}}\,3600\,\frac{\text{sec}}{\text{hr}}\,40\,\frac{\text{ft}}{\text{sec}}} = 16 \text{ pipes [illegible]}$$

$$G = \frac{185\,000\,\frac{\text{ft}}{\text{hr}}}{2 \quad 3600 \quad 16 \quad .0332} = 25.3\,\frac{\#}{\text{sec ft}^2}$$

$$h_{m} = .0225 \frac{k}{D} \left(\frac{D\,G}{\mu}\right)^{.8} \left(\frac{C\mu}{k}\right)^{4}$$

$$= .0225 \frac{.0617}{.206} \left(\frac{.206 \quad 353}{.0275 \quad .000672}\right)^{.8} \left(\frac{9 \quad .0275 \quad 2.4}{.0617}\right)^{.8}$$

$$= 200$$

$$U_1 = \frac{1}{\frac{1}{200} + \frac{2.469}{2.672} \quad \frac{203}{26.12}} \qquad \frac{1}{.005 + .000602} = 178$$

$$T_F = 100 \sqrt[4]{\frac{U_1 + 37.4}{.000492}} = 2570\,°R$$

a heat balance in the radiant section

$$.173 \quad A \left[\left(\frac{T_F}{100}\right)^4 - \left(\frac{T_S}{100}\right)^4\right] + U_c A \left(T_F - 520\right)$$

$$= F_{end}\, 1070 \frac{Btu}{ft^3} - \frac{F\ 4911}{130000}\ 8.45 \quad \left(T_F - 620\right)$$

$$F_{end} = \frac{102 \left(\frac{T_F}{100}\right)^4 + 720\, T_F - 8{,}125{,}000}{12.89 - 32\, T_F} = 81{,}200 \frac{ft^3}{hr}$$

$$Q_{rad} = 102 \left[\left(\frac{T_F}{100}\right)^4 - \left(\frac{T_S}{100}\right)^4\right] = 36{,}850{,}000 \frac{Btu}{hr}$$

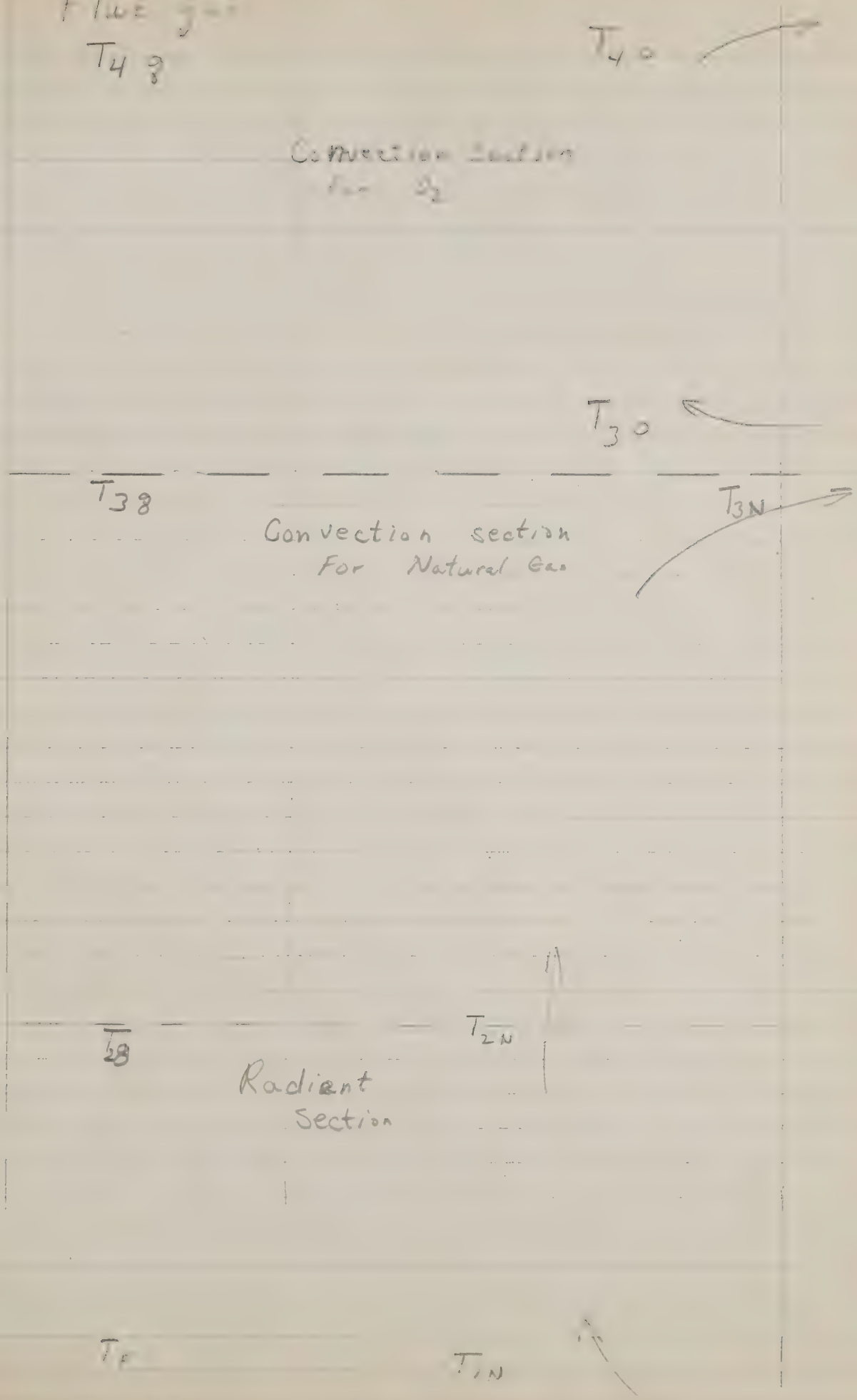
heat balance for the natural gas in rad section

$$Q_{rad} = W_N \int_{T_1}^{T_2} C_p\, dT$$

$$\therefore \quad T_{2N} = 200 \left(-7.5 \pm \sqrt{56.25 + \frac{Q_{rad}}{36000} + 18}\right)$$

$$= 880\,°R$$

$$T_{1g} = T_F - 100 = 2470\,°R$$


Flue gas
T_{4S}
T_{4O}
Convection section
For O_2
T_{3O}
T_{3S}
T_{3N}
Convection section
For Natural Gas
T_{2S}
T_{2N}
Radiant
Section
T_F
T_{1N}

convection section for natural gas

$T_{3N} = 1610\,°K$

$Q_{con} = Q_{total} - Q_{rad} = 5180\ [illegible]$

$c_{p\,avg} = 6.5 + \frac{T_F - 100}{1000} = 6.5 + 2.27 = 8.75\ \frac{[illegible]}{[illegible]}$

A heat balance for the phase gas in the natural gas convection section results in

$T_{3g} = T_F - 100 - \frac{Q_{con}}{.0378\ F\ c_{p\,av}}$

$= 2470 - \frac{15{,}000{,}000}{.0378\ \ 81200\ \ 8.75} = 1910\,°K$

$\Delta T_3 = 1910 - 520 = 1390$

$\Delta T_2 = 2470 - 520 = 1890$

$\Delta T_m = 1490°$

[illegible] 197 [illegible] plot I

$X = \frac{880 - 520}{2470 - 520} = .185$

$Z = \frac{2470 - 1920}{880 - 520} = 1.55$

$\therefore\ Y = 1$

$\Delta T = Y\ \Delta T_m = \Delta T_m = 1490$

$[illegible] = .0225\ \frac{n}{D} \left(\frac{[illegible]}{[illegible]}\right)^2 \left(\frac{[illegible]}{[illegible]}\right)^2$

$= .0225\ \frac{.018}{.206} \left(\frac{[illegible]}{[illegible]}\right)^2 \left(\frac{[illegible]}{[illegible]}\right) = 197$

[illegible]

by [illegible] [illegible] [illegible] [illegible] [illegible]
[illegible] 2 diameters [illegible] [illegible] [illegible]

$h = .108\ G_f^{.562} = .108 \left(F \frac{T_{3p}}{21800}\right)^{.562}$ 16.25

refer to pages 226 and 82 (prob 6) [illegible]

$U_{out} = \frac{1}{\frac{1}{h_{2f}} + \frac{d_o}{k}\frac{d_i}{d_{av}} + \frac{d_i}{D_2 h_{fw}}} = \frac{1}{\frac{1}{16.25} + .0000 + \frac{.125}{[illegible]}}$

$= 15.8$

$Q_{con} = U_{out} A_{out} \Delta T$

$A_{out} = \frac{15\,000\,000}{15.8 \quad 1490} = 636\ ft^2$

L of pipe = $636\ ft^2 \frac{ft^2}{.753\ ft^2} = 845\ ft$ [illegible]

L of pipe [illegible] = 96.5 + 8.16 + [illegible]

= 7460 ft

total length of pipe = 8265 ft

$Re\ No_N = 26800\ G_N = 26800 \quad 35.3 = 945{,}000$

$f = .00026$

$\Delta H = \frac{v^2_{ss}}{16} \frac{L}{D} f = \frac{6800 \quad 8265 \quad .00026}{16 \quad .206} = 4320\ ft$

$\rho = \frac{1874}{359} \frac{300}{14.7} \frac{492}{1033} = .497\ \frac{\#}{ft^3}$

$\Delta P_N = 4320\ ft \quad .497\ \frac{\#}{ft^3} \quad \frac{1\ ft^2}{144\ in^2} = 14.9\ \frac{\#}{in^2}$

Oxygen preheater

a convection section of the natural gas heater

flow 1,[illegible] $\frac{\text{cu ft}^3}{\text{hr}}$

5180 $\frac{\text{moles}}{\text{hr}}$

166,000 $\frac{\text{lb}}{\text{hr}}$

conditions

pressure [illegible]

temp in 710 °R T_{4g}

temp out 1050 °R T_{3g}

$$Q_{co} = 5180 \int_{710}^{1050} (\text{[illegible]} + .001\,T)\,dT = 19000 \text{ [illegible]} \frac{\text{Btu}}{\text{hr}}$$

the Q_{co} for one furnace [illegible] $\frac{\text{Btu}}{\text{hr}}$

$T_{3g} = 1910$ °R

from a heat balance in the convection section of the oxygen heater

$$T_{4g} = T_{3g} - \frac{41900000}{2} = 1490$$

$\Delta T_{3g} = 1910 - 1050 = 860$

$\Delta T_4 = 1490 - 710 = 780$

$\Delta T_m = 820$

$h_{3g} = 1625$

$$S_o = \frac{8330 \frac{\#}{hr}}{3000 \quad .03302 \frac{ft^2}{ft^2} \quad 12 \text{ ft}} = 43.5$$

$$h_{34o} = .3225 \frac{.0212}{.206} \left(\frac{.206 \quad 435}{.03 \quad .000672} \right)^{.8} \left(\frac{.22 \quad .03 \quad 2.42}{.0211} \right)^{.4}$$

$$= 68$$

$$U_{out} = \frac{1}{\frac{1}{1625} + \frac{2.875}{2.469 \quad 12} + \frac{.203 \quad 2.469}{12 \quad 20 \quad 2.870}} = 15.7$$

$$A_{out} = \frac{Q}{U_{out} \, \Delta t} = \frac{6,504,000}{15.7 \quad 860} = 505 \text{ ft}^2$$

$$L = \frac{505}{.753} = 671 \text{ ft}$$

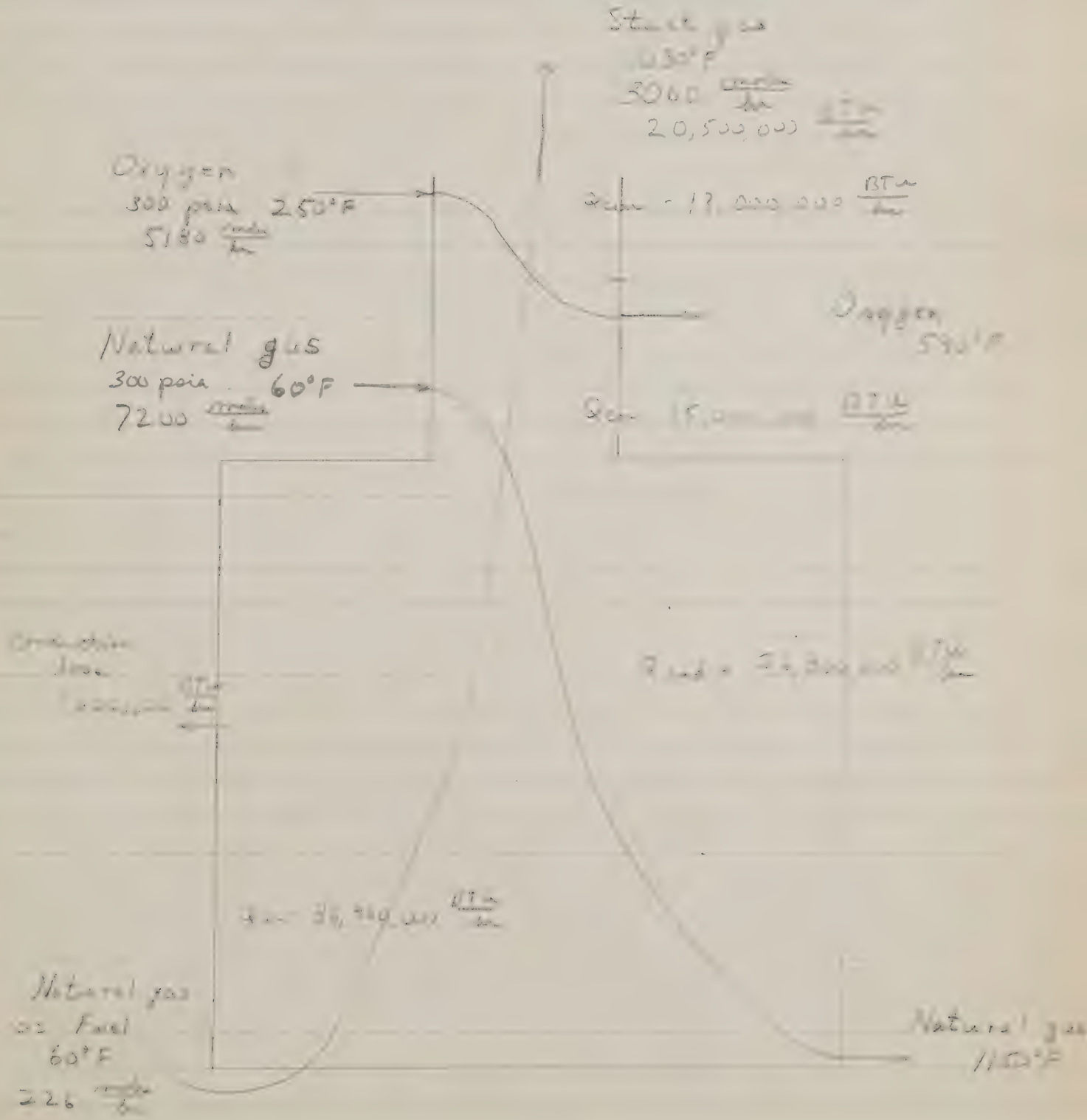
$$h_L N_{Re} = \frac{.206 \quad 43.5}{.03 \quad .000672} = 443000 \qquad f = .00028$$

$$\Delta H = \frac{v^2 L f}{16 D} = \frac{1965 \quad 671 \quad 00028}{16 \quad .206} = 112 \text{ ft}$$

$$\rho = \frac{32}{359} \quad \frac{300}{14.7} \quad \frac{492}{880} = 1.016 \frac{\#}{ft^3}$$

$$\Delta P = \Delta H \rho = \frac{112 \text{ ft} \quad 1.016}{144 \text{ in}^2} \frac{\#}{ft^3} \text{ ft}^2 = .79 \frac{\#}{in^2}$$

Heat Balance



Summary

Calculation made in the radiant section of the preheater followed the method of Tobo and Hottel, with due regard to the influence of the emissivity of the surfaces and of the involved. Allowance was made for the heat loss through the walls, and the temperature in the radiant section was controlled by the overall heat transfer coefficient at the point where the hot natural gas left the furnace. This was necessary because when temperature is fixed at 1200°F, 15% excess air was used in burning.

The arrangement of pipes is also considered with 16 ft pipe spans in the radiant section and 14 ft pipe spans in convection section, which incidentally is placed on top of radiant section. 16-2½" pipes carried the flow throughout the furnace with velocities from 35 ft/sec at the inlet to 110 ft/sec at the outlet for the natural gas. The oxygen flow was also carried in 16-2½" pipes but velocities varied from 35 ft/sec at the inlet to 52 ft/sec at the outlet. The pressure drop in the natural gas section is 15 psi but is only 1 psi in the oxygen section. The difference is due chiefly to the a greater length of pipes in the natural gas section (506 ft as compared to 41 ft for oxygen) and the greater velocities in the natural gas section. Velocities above and below those finally chosen have been considered but both indicate an increased annual cost.

The furnaces are expected to be repaired every 2 years with no salvage value. The total length of pipe used is about 2680 ft and the walls will be of refractory brick 2 ft thick which brings the outside dimensions of the structure to 20 × 30 × 24

Summary of capital investment - [illegible]

Pipe 8935 ft [illegible]

8935 ft × 5.8 $\frac{\text{lb}}{\text{ft}}$ × 1.70 $\frac{\text{lb}}{\text{lb}}$ × 1.50 = 101,000

Brick 5060 ft³

5060 ft³ × 2.95 $\frac{\text{lb}}{\text{ft}^3}$ = 15,000

cost per [illegible] $ 116,000*

total cost [illegible] $ 545,000

Annual cost

[illegible] $ 116,000

heating fuel 162,400 $\frac{\text{ft}^3}{\text{hr}}$ × [illegible]

162,400 $\frac{\text{ft}^3}{\text{hr}}$ × [illegible] = [illegible]

labor 4 operators

Bibliography

A.S.M.E. Power Boiler Code 1942

Chicago, Piping Handbook, 4th Ed.

McAdams, W.H., Heat Transmission 2nd Ed.

Perry, J.H., Chemical Engineers Handbook 3rd Ed.

Synthesis Gas Generator Design

Product Stream		
CO	8260	moles/hr.
H_2	14370	"
CO_2	215	"
H_2O	1150	"
CH_4	23	"
N_2	297	
Total	24335	"
Enthalpy	-327 [illegible] B.T.U./hr	

600°F

Heat taken out by waste heat recovery system 338,200,000 B.T.U/hr

Heat loss thru wall [illegible] B.T.U./hr

1000°F

Feed Stream		
CH_4	6690	moles/hr
C_2H_6	1404	
O_2	4920	
N_2	297	
Total	12311	
Enthalpy	-39[illegible] [illegible] B.T.U./hr	

HEAT AND MATERIAL BALANCE OF SYNTHESIS GAS GENERATOR

Synthesis Gas Generators

I. Introduction

Two methods of synthesis gas production from natural gas have been proposed for use in Fischer - Tropsch plants. They are steam - carbon dioxide reforming (1) and partial oxidation with process oxygen (2). The first method has the advantage that it does not require expensive feed compressors but the reaction is highly endothermic and expensive equipment is required in order to transfer heat to the reaction zone at a high temperature level. The second process can be operated adiabatically. Hence the equipment required is less expensive. But it has the disadvantage of having relatively expensive process oxygen.

In the present plant, partial oxidation of natural gas process is selected because of the following two facts: (1) With the recent developed tonnage oxygen, this process can compete economically with the steam-carbon dioxide reforming where expensive equipment is required; (2) The partial oxidation can be operated at moderately elevated pressure. The Fischer Tropsch reactor of the present plant is designed to operate at 300 p.s.i. Since the synthesis gas generation increases the volume of the gas stream by one fold it is advantageous to operate the synthesis gas generator at around 300 p.s.i.

The partial oxidation of natural gas can be accomplished with or without a catalyst. In the German operation of this process, the products of the initial combustion are passed over a nickel catalyst to achieve equilibrium. In the case of not using a catalyst the temperature level must be in the neighborhood of 2350°F or higher in order for equilibrium to be reached at a reasonable rate. This high temperature level is obtained by using an excess of oxygen, resulting in the formation of carbon dioxide and water vapor at the expense of carbon monoxide and hydrogen. In the present design catalyst is not to be used since there are indications in the literature (2)(3) that in low temperature catalyst operation carbon formation may become a serious problem.

II Thermodynamical calculation

The reactions which take place in the partial oxidation of methane can be summarized by the following four equations

Reaction I. $CH_4 + \frac{1}{2}O_2 \rightleftharpoons CO + 2H_2$

$$K_I = \frac{(CO)(H_2)^2}{(CH_4)(O_2)^{0.5}}\left(\frac{P}{\Sigma N}\right)^{1.5} \qquad (1)$$

Where N represents number of moles and P is pressure in atmosphere

Reaction II $CH_4 + H_2O \rightleftharpoons CO + 3H_2$

$$K_{II} = \frac{(CO)(H_2)^3}{(CH_4)(H_2O)}\left(\frac{P}{\Sigma N}\right)^2 \qquad (2)$$

Reaction III $H_2 + CO_2 \rightleftharpoons H_2O + CO$

$$K_{III} = \frac{(H_2O)(CO)}{(H_2)(CO_2)} \qquad (3)$$

Reaction IV $C + CO_2 \rightleftharpoons 2CO$

$$K_{IV} = \frac{(CO)^2}{(CO_2)}\left(\frac{P}{\Sigma N}\right) \qquad (4)$$

Reaction I can be neglected since the unreacted oxygen is essentially zero at equilibrium. Reaction IV introduces the rate of carbon formation. Since carbon does not enter into the other two reactions, reaction II can be considered separately and leaving only two equations equation (2) and (3). The calculation for equilibrium composition therefore amounts to solving simultaneous equations (2) and (3) with enthalpy and material balance equations.

The feeding natural gas after being scrubbed of H_2S has the following composition.

CH_4	79.8%
C_2H_6	19.7%
N_2	0.5%

The pure oxygen is 95% pure the impurity is N_2 only. From preliminary calculations the amount of CH_4 in the equilibrium mixture is extremely small. Thus the existence of C_2H_6 in the equilibrium mixture can be neglected and equation (1) is used. The enthalpy balance can be written as [neglecting C_2H_6 unreacted]

$$N_{O_2}H_{O_2} + N_{CH_4}H_{CH_4} + N_{C_2H_6}H_{C_2H_6} + N_{N_2}H_{N_2} = N'_{CO}H'_{CO} + N'_{H_2}H'_{H_2} + N'_{H_2O}H'_{H_2O} + N'_{CO_2}H'_{CO_2} + N'_{N_2}H'_{N_2} + N'_{CH_4}H'_{CH_4} + Q \qquad (5)$$

where the N and H's are the initial number of moles and initial enthalpies respectively, N' and H'$_i$'s are the number of moles and total enthalpies at equilibrium. Q is the amount of heat lost through the reactor wall.

The material balances are

O_2 balance: $$N_{O_2} = \frac{N'_{H_2O}}{2} + N'_{CO_2} + \frac{N'_{CO}}{2} \qquad (6)$$

C balance. $$N_{CH_4} + 2N_{C_2H_6} = N'_{CO} + N'_{CO_2} + N'_{CH_4} \qquad (7)$$

H_2 balance: $$2N_{CH_4} + 3N_{C_2H_6} = N'_{H_2} + N'_{H_2O} + 2N'_{CH_4} \qquad (8)$$

N_2 balance: $$N_{N_2} = N'_{N_2} \qquad (9)$$

From feed composition we have.

$$N_{N_2} = N_{CH_4} \times \frac{0.5}{79.8} + N_{O_2} \times \frac{5}{95} \qquad (10)$$

$$N_{C_2H_6} = N_{CH_4} \times \frac{19.7}{79.8} \qquad (11)$$

There are nine independent equations equation (2)(3)(5)(6)(9)(10)(11) and eleven variables. By assuming $N_{CO} = 100$ and Q = 4,800 B.T.U. The equilibrium mixture can be solved.

In solving these equations equilibrium temperature is chosen at 2400°F. The feed gases are preheated to 1000°F and pressure is at 20 atmospheres. The enthalpies in consistent units are from the extrapolation of the data in reference (4), the enthalpy of C_2H_6 is calculated from its heat of formation and heat capacities. The equilibrium constants are also extrapolated values. Table I are the enthalpies. Table II are the equilibrium constants.

Table I. Enthalpies of the Gases (in B.T.U./moles)

Temperature	O_2	N_2	H_2	CH_4	C_2H_6	H_2O	CO	CO_2
600°F	—	7458	7,287	-19,002	—	-94,360	-44,[illegible]	-159,697
1000°F	10,690	10,345	—	-13,492	-13,642	—	—	—
2400°F		21,630	20,460	12,400	-	-76,000	-27,107	-136,000

Table II.

	K_I	K_{II}	K_{III}	K_{IV}
2400°F	—	6.60×10^5	3.08	—

By substituting these quantities into equations (2)(4)(5)(6)(7), we have the following set of equations

$$\frac{100\, N'^3_{H_2}}{N'_{CH_4}\, N'_{H_2O}} \left[\frac{20}{100 + N'_{H_2} + N'_{H_2O} + N'_{CH_4}} \right]^2 = 6.60 \times 10^5 \qquad (12)$$

$$\frac{100\, N'_{H_2O}}{N'_{H_2}\, N'_{CO_2}} = 3.08 \qquad (13)$$

$$10{,}690\, N_{O_2} - 13492\, N_{CH_4} - 13642\, N_{C_2H_6} + 10345\, N_{N_2}$$

$$= -27{,}100 \times 100 + 20{,}460\, N'_{H_2} - 76{,}000\, N'_{H_2O} - 136{,}000\, N'_{CO_2} + 21{,}630\, N'_{H_2} + 12{,}400\, N'_{CH_4} + 4{,}800 \qquad (14)$$

$$N_{O_2} = \frac{N'_{H_2O}}{2} + N'_{CO_2} + 50 \qquad (15)$$

$$N_{CH_4} + 2N_{C_2H_6} = 100 + N'_{CO_2} + N'_{CH_4} \qquad (16)$$

Solving equations (8) to (16) we have

$N_{O_2} = 59.5$	$N'_{CO} = 100$
$N_{CH_4} = 68.9$	$N'_{H_2} = 174$
$N_{C_2H_6} = 17.0$	$N'_{CO_2} = 2.6$
$N_{N_2} = 3.6$	$N'_{H_2O} = 13.9$
	$N'_{CH_4} = 0.27$
	$N'_{N_2} = 3.6$

Thus under the present condition the ratio of natural gas to process oxygen in the feed is 1.4 and the ratio of H_2 and CO at equilibrium is 1.74

The synthesis gas composition required by the Fischer Tropsch reactor is H_2 to CO at a ratio of 1.9 and at a rate of 190,000,000 standard cubic foot per day. The H_2, CO ratio is raised by passing a portion of the stream through a water shift reactor. Also 500 moles per hour of pure H_2 is to be provided for the reduction of iron catalyst of the Fischer Tropsch

reactor. From overall material balance the amount of CO required is 8260.5 moles/hr. Putting up the feed stream and the product stream we have

Feed Stream at 1000°F

CH_4	5690	moles/hr.
C_2H_6	1404	"
O_2	4920	"
N_2	297	"
Total	12311	"
Total enthalpy	−39,900,000 B.T.U./hr.	

Product Stream at 2400°F

CO	8260.5	moles/hr
H_2	1439.0	"
CO_2	215	"
H_2O	1150	"
CH_4	22.5	"
N_2	297	"
Total	12334	"
Total enthalpy	−40,300,000 B.T.U./hr.	

Amount of heat allowable to lose through reactor wall = 40,300,000 − 39,900,000 = 400,000 B.T.U./hr

The product leaving the reaction zone must be cooled as fast as possible to approximate 1000°F or lower in order to stop the reaction and to retain the 2400°F equilibrium composition. In the Texaco process this is done by quenching the gas stream with water. Since this plant is operating at 20 atm the gas side film coefficient of heat transfer is higher than that at ordinary pressure. This heat therefore is utilized in a waste heat recovery system to generate steam at 680 p.s.i. This waste heat recovery system consists mainly of a bank of boiler tubes. It is located inside the gas generator just above the reaction zone. The synthesis gas is to be cooled to 600°F by this system.

The total enthalpy of the synthesis gas at 600°F has a total enthalpy of −378,500,000 B.T.U./hr (Calculated from enthalpy data in table I)

∴ Total amount of heat to be removed = −40,300,000 + 378,500,000 = 338,200,000 B.T.U./hr

The heat required to generate one pound of 600 p.s.i. steam assuming that feed water enters at 85°F is (the saturation temperature of 600 p.s.i. steam is 500°F)

$$500 - 85 + 714 = 1129 \text{ B.T.U.}$$

Assuming that the waste heat recovery system is of 90% efficiency, then the total amount of steam generated is

$$\frac{338,200,000}{1129} \times \frac{90}{100} = 270,000 \text{ lb/hr.}$$

Feed water required is

$$\frac{270,000 \times 7.481}{60 \times 62.4} = 540 \text{ gal/min.}$$

II Design and Cost

A) Contact time. The contact time required for conducting the process at pressures above 200 p.s.i. is about three to four times the contact time at atmospheric pressure. In general it will fall in the range of about 1 to 20 seconds and preferably 10 seconds (3). No reaction rate data of this process are found in literature except for low temperature operation using catalyst (4). If activation energy at such low temperature is calculated and used to estimate the reaction rate for present conditions a very large extent of uncertainty will involve. However, in reference (3) the author gave an optimum contact time of 6 sec for this reaction at 250 p.s.i.g. and at 2500°F. Hence a contact time doubling this figure namely 12 sec. is chosen for the present condition 500 psi and 2400°F.

B). Gas Velocity in the reaction zone

The limiting factor for reacting gas velocity is the flame velocity of natural gas in oxygen. The velocity of the gases in the reaction zone should be smaller than flame velocity so that combustion is possible.

Direct information on flame velocity under the present condition, namely the burning of methane-steam mixture in 95% O_2-N_2 mixture preheated to 1000°F and at a pressure of 300 p.s.i.a., has not been found in the literature. From Perry's Handbook 3rd ed. p. 1587 the flame velocity of 20% natural gas in air (corresponds to 1.4 natural gas to O_2 ratio) is 0.5 ft/sec. Jahn (5) showed that the flame velocity of H_2 in 95% O_2 is approximately 4.5 times as fast as that in air. According to Ubbelohde (6) (7) preheating of the gases increases the flame velocity moderately while increase of pressure decreases the flame velocity moderately. For the present case the effects of temperature and pressure on flame velocity can be assumed compensating each other. So the flame in laminar flow is approximately $4.5 \times 0.2 = 0.9$ ft/sec. If a turbulence factor of 20 is taken, the flame velocity is then 18 ft/sec. This 18 ft/sec is used as an upper limit for the gas velocity in designing the reactor.

Since the contact time is 12 sec., a velocity of 5 ft/sec will require a reaction zone 60 ft high. Together with the space needed for cooling system, foundation etc. the height of the generator will be close to 100 ft. Any velocity higher than 5 ft/sec is then not practical. The optimum velocity as it is shown later from economical balance is 2 ft/sec.

c) Design Calculations:

The shape of the synthesis gas generator is shown in the attached figure. The size of the generator, the length of the boiler pipes in the waste heat recovery section and the thickness of refractory lining depend on the number of units to be used and the velocity of gases in the reaction zone. One spare unit is always necessary since the change of refractory lining and cleaning of boiler tubes require temporary shut down. An economic balance is tabulated in table III.

Table IV

Gas velocity in ft/sec	No. of units required	Inside diameter of the outer shell (& the outside diameter of refractory) in ft	Total Height in ft	Cost per year of steel, refractory lining and boiler pipes
5	2	14.8	89	21060
	3	12.4	89	29860
3.5	2	16.3	68	18630
	3	13.6	68	26350
2	2	19.4	47	18070
	3	15.6	47	22300
1	2	23.8	33	22800
	3	19.3	33	22450
	4	16.8	33	26900

The calculations for each of the cases in table III are the same as the step for the optimum case, 2 units with gas velocity of 2 ft/sec, as is shown in the following section.

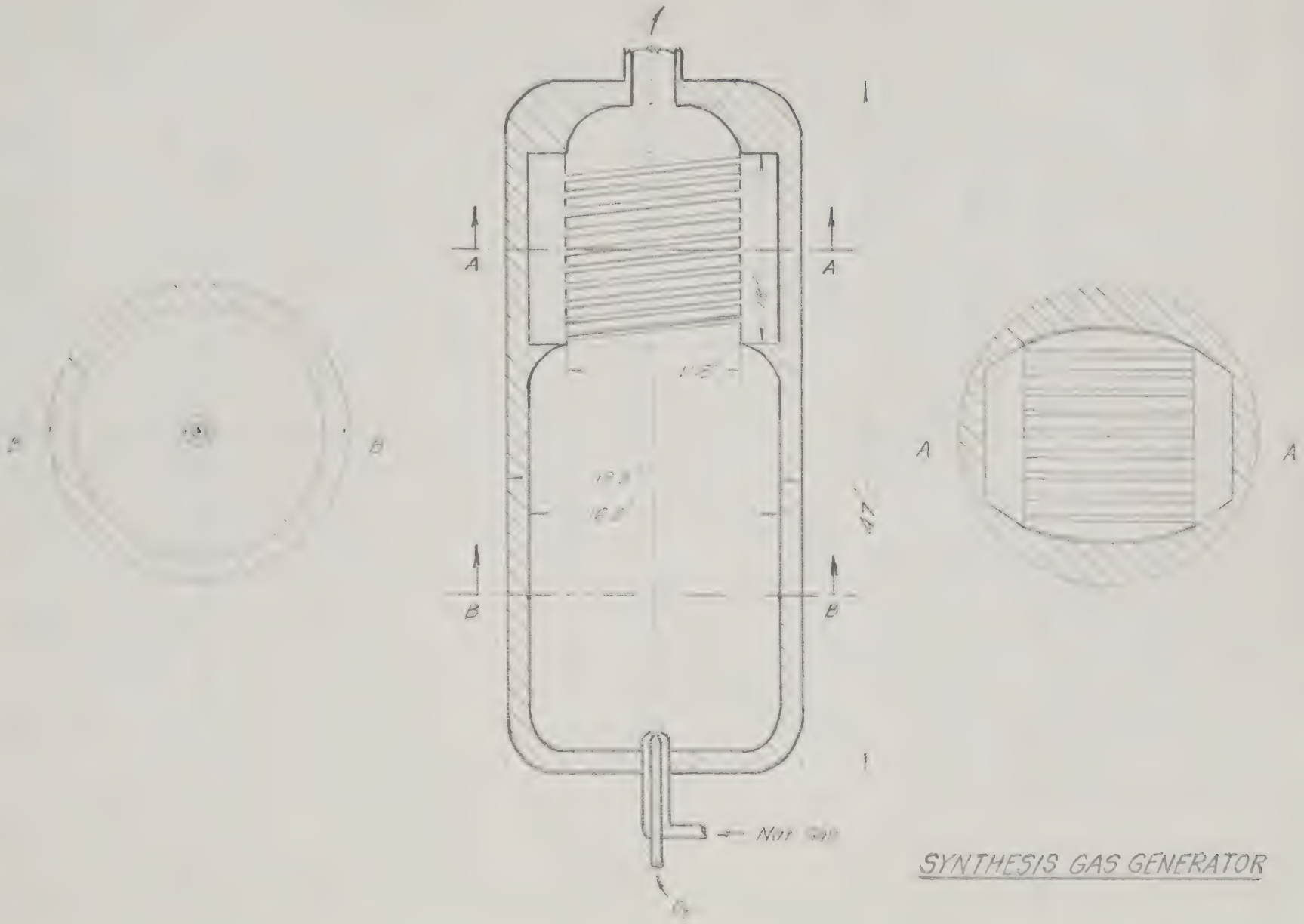
Av. gas velocity = 2 ft/sec

Feed gas volume $= 12{,}311 \times 359 \times \frac{1}{20} \times \frac{1460}{492} = 656{,}000$ ft³/hr

Product gas volume $= 24{,}335 \times 359 \times \frac{1}{20} \times \frac{2460}{492} = 2{,}540{,}000$ ft³/hr

Average volume = 1,598,000 ft³/hr

or = 444 ft/sec



SYNTHESIS GAS GENERATOR

Cross sectional area $= \frac{444}{2} = 222\ ft^2$

Reaction zone height $= 2 \times 12 = 24\ ft$

Inside diameter of the reactor $= \sqrt{\frac{4 \times 222}{\pi}} = 16.8\ ft$

2 inches o.d. boiler tubes are used in waste heat recovery section. These are arranged in equilateral triangles and spaced to centre distance of 4 inches

Length of boiler tube per section $= \sqrt{\left(\frac{16.8}{2}\right)^2 + \left(\frac{16.8}{2}\right)^2} = 11.88\ ft$

Number of tubes in one row $= \frac{11.88}{\frac{4}{12}} = 36$

Minimum free cross section area in the tube bank $= \frac{222}{2} = 111\ ft^2$

Mass rate of flow $= 4920 \times 32 + 297 \times 28 + 1404 \times 30 + 5690 \times 16$
$= 298,740\ lb/hr.$

Maximum mass rate of flow per unit area $= \frac{298,740}{111} = 2690\ lb/(hr)(ft^2)$

The gas side film coefficient of heat transfer is calculated by means of Monrad's equation on page 472 Perry's Handbook 3rd ed

$$h_m = \frac{0.031\, C_p\, T^{0.3}\, G_{max}^{2/3}}{(D_o')^{1/3}} \qquad (7)$$

Where T is the average gas temperature in degree Rankine

$$\left[T = \frac{2400 + 600}{2} + 460 = 1960°R\right]$$

G_{max} is the maximum mass rate of flow per unit area of the gas
D_o' is the outside diameter of the boiler tubes in inches
C_p is the weight average of the specific heats of the components in the gas phase

At 1960°R $\quad C_{p_{H_2}} = 3.5 \quad C_{p_{CO}} = 0.28 \quad C_{p_{H_2O}} = 0.55$

$$C_{p_{gas}} = \frac{8260.5 \times 28 \times 0.28 + 1639.0 \times 2 \times 3.5 + 1150 \times 0.55 \times 18}{8260.5 \times 28 + 1639.0 \times 2 + 1150 \times 18} = 1.19\ B.T.U./lb.$$

$$\therefore h_m = \frac{0.031 \times 1.19 \times 1960^{0.3} \times 2690^{2/3}}{2^{1/3}} = 56.5\ B.T.U./(hr)(°F)(ft^2)$$

The water side film coefficient of heat transfer and the conductivity of the tube walls are large compared to the gas film coefficient (8). The overall coefficient U can therefore be approximated by using the value of h_m.

The logarithmic mean temperature drop across the tube wall is

$$\Delta T_m = \frac{(2400-500)-(600-500)}{2.3 \log \frac{2400-500}{600-500}} = 613°F$$

The area of boiler tube required can be calculated from

$$A = \frac{Q}{U \Delta T_m} = \frac{338,200,000}{613 \times 55.5} = 9950 \text{ ft}^2$$

Total length of boiler tube required for one unit is

$$\frac{9950}{\frac{2}{12} \times \pi} = 19,000 \text{ ft}$$

Since one spare unit is required the total length of tube required is

$$2 \times 19,000 = 38,000 \text{ ft.}$$

Number of rows of boiler tube needed is

$$\frac{19,000}{36 \times 11.9} = 45 \text{ rows}$$

The distance between two rows is

$$\sqrt{4^2 - 2^2} = \sqrt{12} = 3.46 \text{ inches}$$

Hence the height of the tube bank is

$$45 \times \frac{3.46}{12} = 13 \text{ ft}$$

Conductivity of insulation bricks at 2500°F around is 0.183 (B.T.U.)/(hr)(ft)(ft)

Gas film coefficient of heat transfer inside the reactor wall can be calculated by equation (18) [equation (24) on page 467 Perry's Handbook 3rd ed.]

$$h = \frac{16.6 C_p (G')^{0.8}}{(D_i')^{0.2}} \qquad (18)$$

C_{pav} at 2400°F is

$$\frac{8260.5 \times 28 \times 1.20 + 14390 \times 2 \times 3.6 + 1150 \times 0.62 \times 18}{72580} = 1.28$$

Mass rate of flow per unit area at reaction zone is

$$G' = \frac{298,740}{262 \times 3600} = 0.374 \text{ lb/(ft}^2\text{)(sec)}$$

Hence

$$h = \frac{16.6 \times 1.28 \times (0.374)^{0.8}}{(16.8 \times 12)^{0.2}} = \frac{16.6 \times 1.28 \times 0.455}{2.89} = 3.35 \frac{\text{B.T.U.}}{(\text{hr})(°\text{F})(\text{ft}^2)}$$

Allowable heat loss through reactor wall is 400,000 B.T.U/hr

Assume that the temperature of the steel shell is at 100°F, then

$$\Delta t = 2400 - 100 = 2300°\text{F}$$

Hence

$$U = \frac{400,000}{16.8 \times 24 \times \pi \times 2300} = 0.137 \frac{\text{B.T.U.}}{(°\text{F})(\text{hr})(\text{ft}^2)}$$

Since

$$\frac{1}{U} = \frac{1}{h} + \frac{x}{k} \qquad (19)$$

where x is the thickness of insulation brick and k is the thermal conductivity of the insulation brick

$$x = \left(\frac{1}{0.137} - \frac{1}{3.35}\right) 0.189 = 1.28 \text{ ft}$$

The inner diameter of the steel shell is then

$$16.8 + 2 \times 1.28 = 19.4 \text{ ft}$$

The thickness of the steel shell can be calculated by means of equation (20) [equation (1) page 1238 Perry's Handbook 3rd ed.]

$$t = \frac{PD}{2S} \qquad (20)$$

The ultimate strength of plain carbon steel is 60,000 lb/sq. in., if a safety factor of 5 is used

$$S = \frac{60,000}{5} = 12,000 \text{ lb/sq. in}$$

$$t = \frac{500 \times 19.4 \times 12}{2 \times 12000} = 2.91 \text{ inch thick}$$

The outside diameter of the reactor is therefore $19.4 + \frac{2.91 \times 2}{12} = 19.9$ ft.

If 7.5 ft is allowed for the clearance between the tube bank and the gas outlet and between the reaction zone and gas inlet the total height of the reactor is

$$24 + 13 + 7.5 + 2 \times 1.28 \cong 47 \text{ ft}.$$

Total amount of steel required is then (including one spare unit) =

$$\frac{2.91}{12}\left[19.4 \times \pi \times 47 + 2 \times \pi \times \left(\frac{19.4}{2}\right)^2\right] \times 7.8 \times 6.24 \times 2 = 815{,}000 \text{ lb}.$$

Total volume of insulation brick required (including one spare unit) is

$$\left\{\left[\left(\frac{19.4}{2}\right)^2 - \left(\frac{16.8}{2}\right)^2\right] \times \pi \times 47 + \left(\frac{19.4}{2}\right)^2 \times \pi \times 2 \times 1.28\right\} \times 2 = 5{,}074 \text{ ft}^3$$

D) Pressure drop of the gases

Pressure drop due to the bank of tubes can be calculated by means of equation (21)

$$\Delta P = 4 \rho f''' N V_{max}^2 / 2g_c \qquad (21)$$

where N is number of rows in the bank, V_{max} is the maximum velocity in ft/sec and

$$f''' = \left(0.23 + \frac{0.11}{(X_t - 1)^{1.08}}\right)\left(\frac{D_o G_{max}}{\mu_f}\right) \qquad (22)$$

For bank of staggered tubes where X_t = pitch/D_o = 2, equation (22) reduces to

$$f''' = 0.34\left(\frac{D_o G_{max}}{\mu_f}\right)^{-0.15} \qquad (23)$$

The film temperature can be approximated by

$$\frac{1500 + 500}{2} = 1000°F$$

For CO $\mu_f = 0.035$ centipoise
For H_2O $\mu_f = 0.030$ "
For H_2 $\mu_f = 0.018$ "

If the largest value namely 0.035 centipoise is used for the gas mixture

$$f''' = 0.34 \left(\frac{\frac{2}{12} \times 2690}{0.035 \times 2.42} \right)^{-0.15} = \frac{0.34}{3.62}$$

$$V_{max} = \frac{2,540,000}{111 \times 3600} = 6.35 \text{ ft/sec}$$

$$\rho = \frac{298,740}{2,540,000} = 0.118 \text{ lb/ft}^3$$

Hence $$\Delta P = 4 \times \frac{0.34}{3.62} \times 45 \times \frac{(6.35)^2}{2 \times 32.2} \times 0.118 \times \frac{1}{144} = 0.0086 \text{ lb/sq. in}$$

The pressure drop through the gas generator is therefore negligible

E) Cost Estimation

The cost of steam including fabrication and erection is \$0.099 per lb (9). The initial cost of steel is then

$$815,000 \times 0.099 = \$ 80,500$$

The cost of boiler tube is estimated to be \$1.20 per foot. The initial cost for boiler tube is then

$$38,000 \times 1.20 = \$ 45,600$$

Boiler accessories were \$35 for 250 p.s.i.a. boiler per 100 sq ft in 1951 (9). A factor of 3 is used for the present 680 p.s.i.a. boiler. The cost data are based on ordinary boiler where overall heat transfer coefficient is at 13 to 15 (B.T.U.)/(ft²)(°F)(hr). For the present case therefore a factor of $\frac{55.5}{13}$ should be applied. Also a factor of 160/70 should be applied to give the cost in 1950 (10). Hence the cost of boiler accessories is

$$35 \times 3 \times \frac{160}{70} \times \frac{55.5}{13} \times \frac{9950 \times 2}{100} = \$ 204,500$$

The cost of labor for boiler erection is estimated to be \$ 17,630

The amortization of above items is assumed to be 10 years

The cost of insulation bricks is estimated to be \$86 per 1000 pieces of standard size bricks. erection cost is 100%, which gives cost per ft³ to be

$$\frac{86 \times 2}{1000 \times \frac{9}{12} \times \frac{4.5}{12} \times \frac{2.5}{12}} = \$\ 2.93 \text{ per } ft^3$$

Allow 10% extra for special shape bricks required in the roof. The total cost of insulation bricks is

$$2.93 \times \frac{110}{100} \times 8474 = \$\ 27,300$$

Ordinary refractory bricks can last about 2½ years of operation. Since in the present design two units are used alternatively the amortization can be taken as five years.

Hence total initial cost is

$$80,500 + 45,600 + 204,500 + 17,630 + 27,300 = \$371,500$$

Total cost per year is

$$(80,500 + 45,600 + 204,500 + 17,630)\frac{1}{10} + 27,300\,\frac{1}{5} = \$\ 39,660$$

IV Bibliography

(1) Peck, E.B. and Evans, L.F.; U.S. Bur. Mines Inform. Circ. 7375 (August 1946) Edited by R. Holroyd.

(2) Hall, R.E. and Smith, J.M.; Chem. Eng. Progr. 45 459 (1949).

(3) Riblett, E.W.; U.S. Patent 2,491,518 Dec. (1949).

(4) Eichner, C., Perrin, M. and Prettre, M.; Compt. Rend., 48 621, (1944)

(5) Jahn, G.; Karlsruhe Dissertation (1934).

(6) Ubbelohde, L. and Koelliker, E.; Gas-u. Wasserfach, 59, 47-49 (1916)

(7) Ubbelohde, L. and Hofsäss, M., ibid, 56, 1226, 1253 (1913)

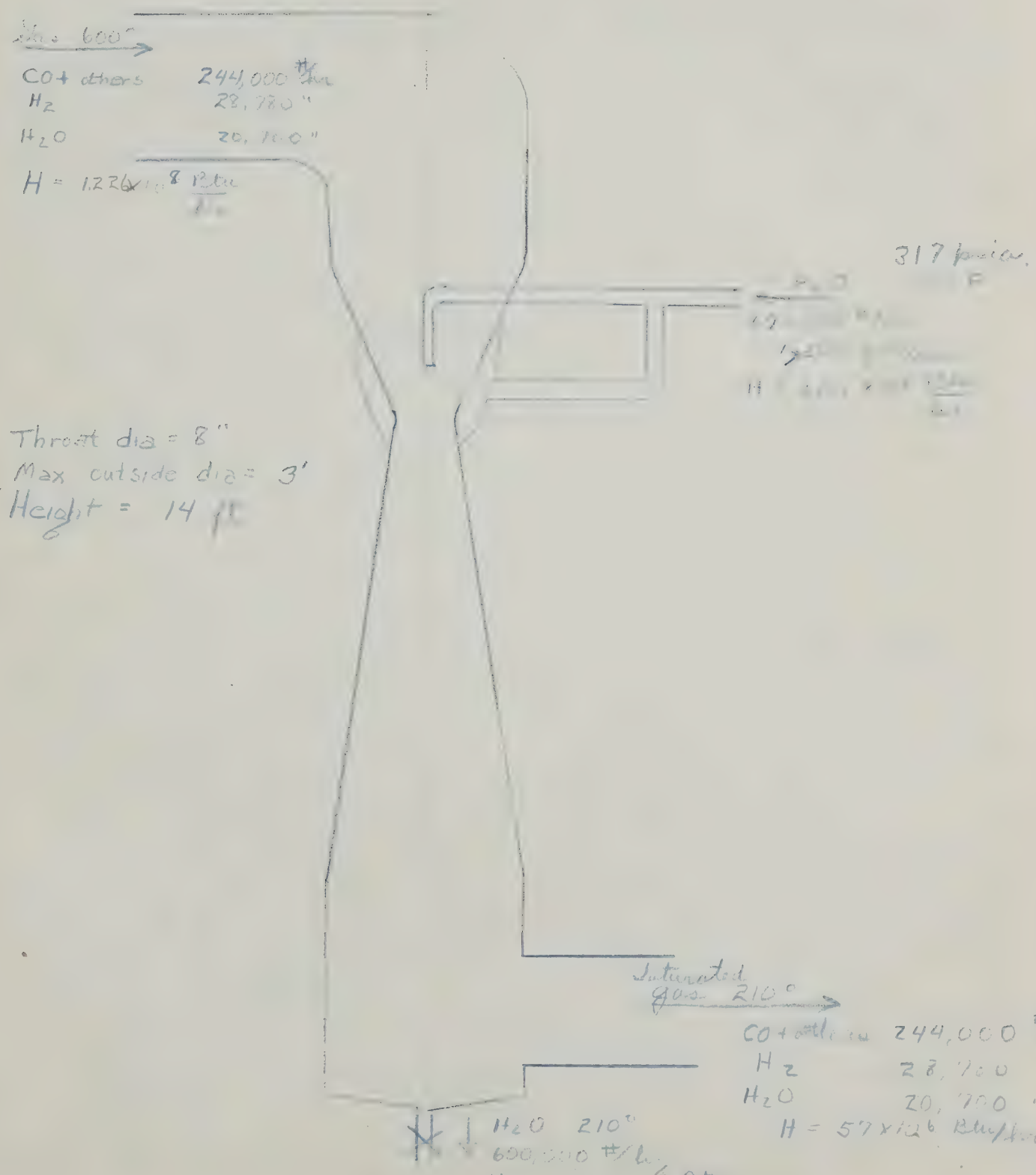
(8) Straub, F.G.; Private communication

(9) Vilbrandt, F.C.; "Chemical Engineering Plant Design" 3rd Ed. McGraw-Hill Book Company, Inc. New York (1949).

(10) Chem Eng. 57 123. (1950)

= Synthesis Gas Scrubber Design =

Synthesis Gas Scrubber #1



Water Water Scrubber

The gas comes from the spillway gas producer at heat boiler at 600°F and so the [illegible] particles of carbon, which are formed [illegible] fine [illegible] and [illegible] the scrubber [illegible] to remove the [illegible] dust, and to cool the gas down to 250° at which temperature the scrubber with water [illegible] a low enough to give a product gas containing only 5% water vapor.

[illegible] the sensible heat of the water only, which [illegible] to cool the 293,000 # of entering gas from 600°F to 250°F. [illegible]

A graph and [illegible] permit the calculation of the [illegible] that the [illegible] a water scrubber would give were given from a [illegible]

[illegible]

Heat balance. Basis 1 hr.

$$24{,}000\,(600°F - 210°F)\left(0.25\,\frac{Btu}{\#\,°F}\right) + 28{,}000\,(600 - 210°F)\left(3.4\,\frac{Btu}{\#\,°F}\right)$$

$$+\ 20{,}700\,(1335 - 1150) = W\,(\ldots)$$

$$W = 6 \times 10^5\ \#\,H_2O/hr$$

594,000 cu ft/hr.

[illegible]

[illegible] 10 [illegible]

[illegible] the [illegible] about 10.5 [illegible] [illegible] [illegible] [illegible] [illegible] [illegible] of the [illegible] Tube.

[illegible] 25° [illegible] of the [illegible] 3 [illegible]

No. [illegible]

$$\frac{1200 \text{ gal/min}}{60 \times 7.48 \times \left(\frac{3}{8}\right)^{[illegible]}} = [illegible]$$

[illegible]

30 [illegible] every 1" [illegible] [illegible]

This amounts to 1 [illegible]

[illegible]

[illegible]

[illegible]

[illegible]

[illegible]

[illegible]

[illegible] pressure drop [illegible]

[illegible] loss at throat is never resumed [illegible]

[illegible]

$$\frac{100 \times 144}{.29} + \frac{(25)^2}{64.3} + 144 \times 300\left(\frac{1}{.45} - \frac{1}{.29}\right) = \frac{(250)^2}{64.3} + \frac{p \times 144}{.45}$$

$p = 297$ psi

[illegible]

Calculation of rigid water pressure

[illegible]

[illegible] for 8 pipe [illegible] psi/ft

[illegible] to throat L [illegible]

[illegible] 1 psi drop [illegible]

Hence, [illegible] must be compressed to 31 psi

Cost

# of steel pressure = $14 \times \frac{25}{18} \pi \times \frac{0.5}{12} \times 7.8 \times 62.4 = 25.55$#

[illegible] steel [illegible]

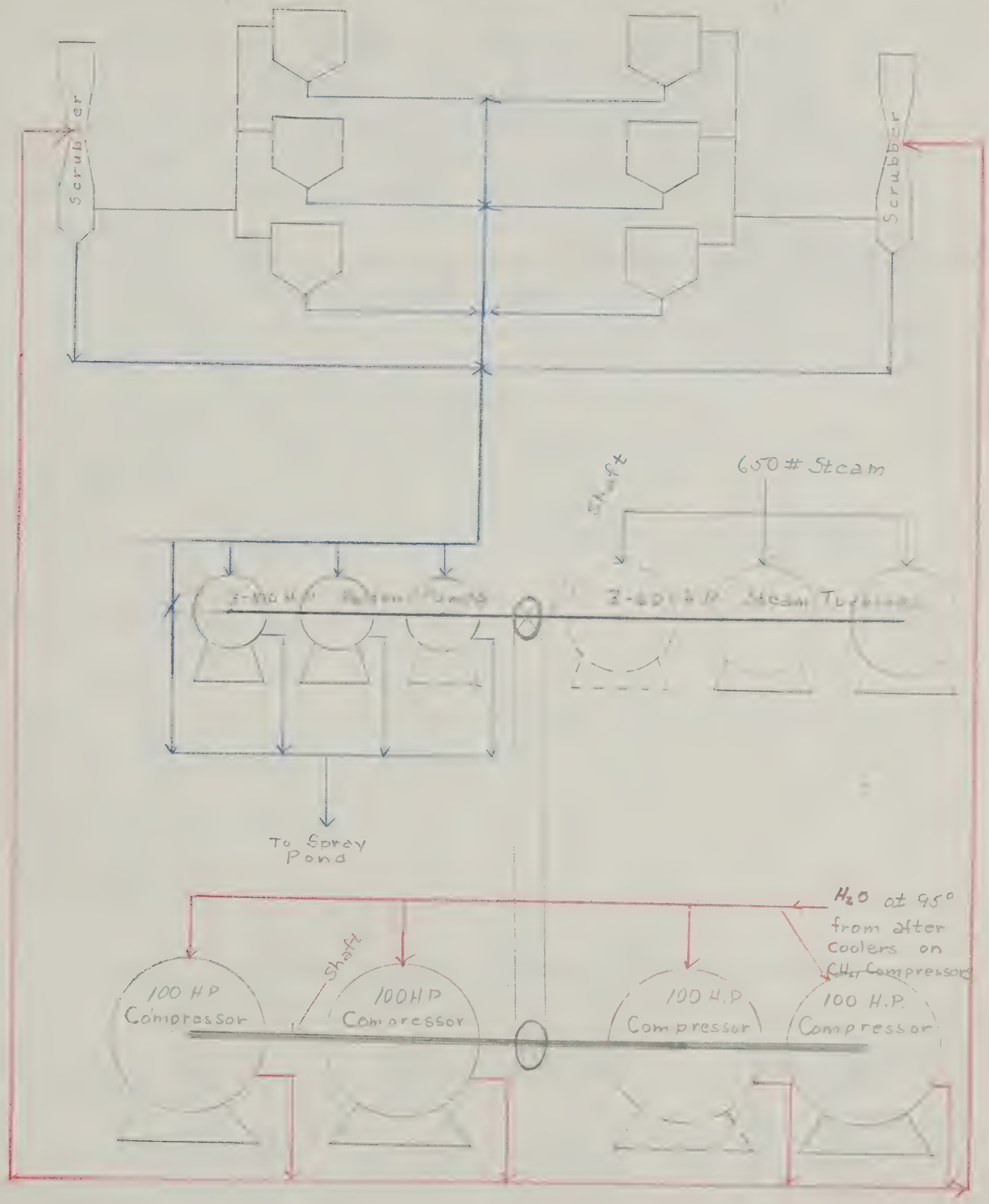
[illegible]

[illegible]

[illegible]

[illegible]

Outline of Compressor Plant, Compressing Water to #1 Scrubber



— Hot water

— Cold Water

[illegible] water to #1 scrubber

1200 [illegible] of water must be compressed to 317 psia in order to force it through the nozzles in the venturi scrubber. [illegible] the water comes out at 245 psia [illegible] to make use of this pressure to [illegible] the [illegible] gas.

To [illegible] the pressure work, an impulse Pelton generator can be used. [illegible] cost [illegible] on this piece of equipment, so conservative values were assumed. Pilot plant data should be taken to make a more accurate design.

[illegible]

$$(245 - 15\text{ psi}) \times \frac{34}{15\text{ psi}} = [illegible]$$

$$\frac{[illegible]}{33{,}000} = [illegible]$$

The water will be [illegible] effective [illegible]

[illegible] horse power is

$$(317 - 15) \times \frac{34}{15} \times \frac{1{,}200 \times 7.33}{33{,}000 \times 0.70} = 297 \text{ H.P.}$$

Horse power necessary from steam turbines is
297 − 173 = 124 H.P.

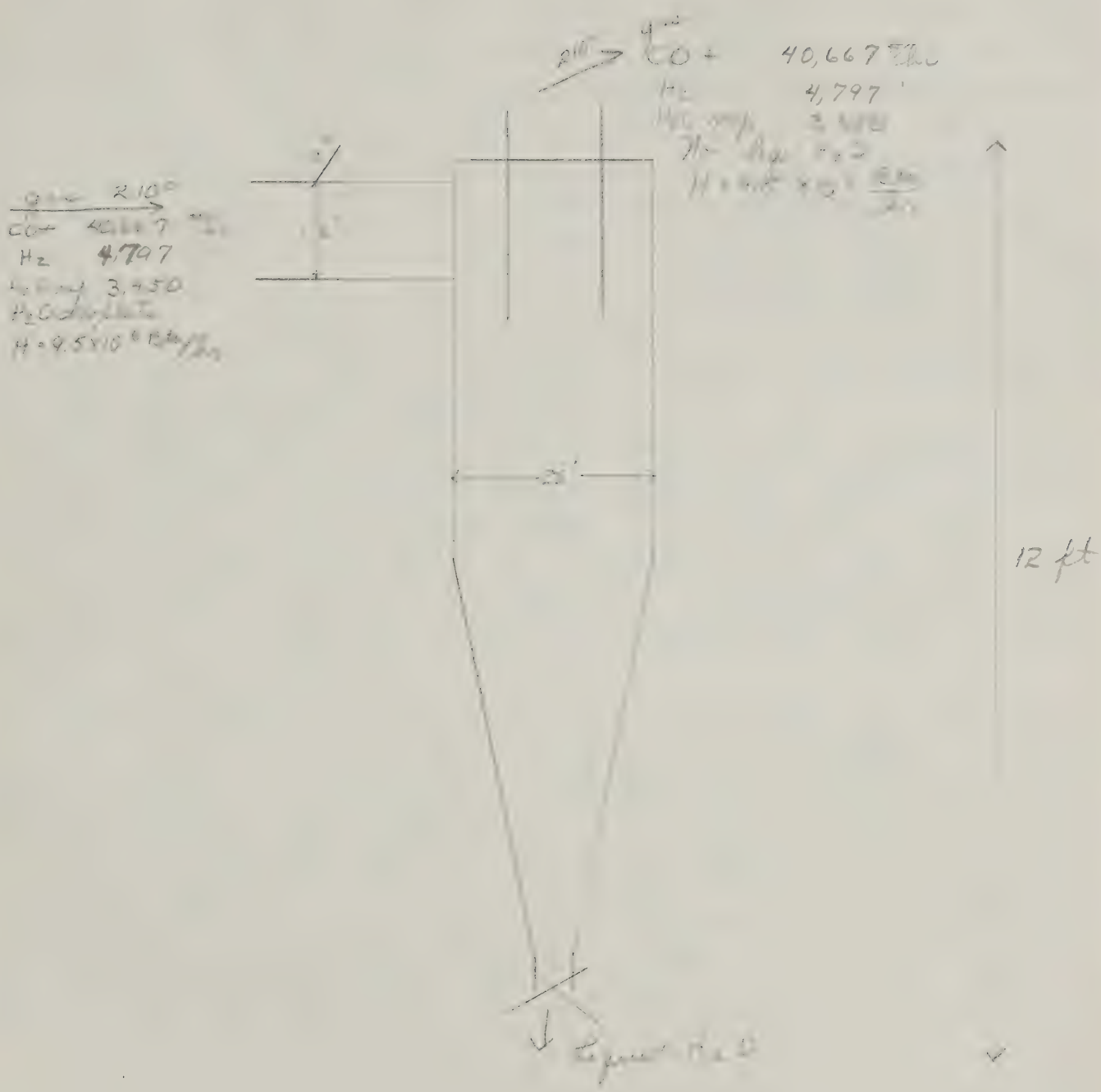
[illegible] $\frac{124 \times 33{,}000 \times 60}{778 \times 270} = 1170 \frac{\#\text{ steam}}{\text{hr}}$

[illegible]

Cost of [illegible] at [illegible]
Cost of Pelton generator at [illegible]
Cost of steam turbines at [illegible]

[illegible] = [illegible]

Cyclone #1 6 of the



Cyclone #1 - for Synthesis Gas

The gas leaving the venturi scrubber must pass through a mitered elbow which will remove most of the water. However some appreciable water [illegible] will pass through this elbow and it is necessary that these [illegible] water [illegible] before the gas goes to the Reactor. [illegible]
Due to the high water flow rate and the velocity and the [illegible] the water droplets will be fairly large, and assuming that [illegible] [illegible] drops greater than [illegible] diameter should [illegible] on the [illegible].

The problem then is to remove these drops. [illegible] [illegible] particles below [illegible] and [illegible] [illegible] [illegible]. Due to the [illegible] ([illegible] ft³/sec) it is [illegible] that any scheme [illegible] centrifugal force would be [illegible].

Design factors + equations were [illegible] [illegible] and it was based on 45¢/lb steel plus $500 per cyclone miscellaneous costs. [illegible] showed that the optimum number and size [illegible] 2 cyclones [illegible] 2 ft in diameter. [illegible] [illegible] [illegible] [illegible] [illegible] could [illegible] [illegible] the [illegible].

[illegible] design [illegible]

Feed	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]
H_2O [illegible]	[illegible]	[illegible]
	[illegible]	[illegible]

[illegible] 210° & about 295 psi

[illegible] = 606 000 [illegible]

(1) [illegible] 2 cyclones are used.

$$\text{Vel} = V = \frac{303\,000}{[illegible]} = \frac{303\,000 \times 3}{D_c^2 \times 3000}$$

[illegible]

$$d_{min} = \sqrt{\frac{9\mu\,(\rho_2 - \rho_1)}{\pi\, V_2\, N\, \rho_s - (\rho)}}$$

[illegible]

$$[illegible] = 3\sqrt{\frac{[illegible]}{3 \times 3.14 \times 30.0 \times \frac{674}{D^2} \times [illegible]}}$$

D_{max} = [illegible] ft

[illegible]

[illegible]

Thickness [illegible] $= \frac{D}{4}\left(\sqrt{\frac{[illegible]}{[illegible]}} - 1\right) = \frac{[illegible]}{[illegible]} D = [illegible]$

Area of walls $\cong 6D(\pi D) = 19D^2$

[illegible] $D(19D^2)(7.8 \times 62.4)(\$0.45) = \$54.2D^3$

Cost of valves [illegible] + V.C. $= 54.2D^3 + 300$

[illegible]

$[illegible] = \frac{[illegible]}{D_e^2}$ [illegible] 1024

$= [illegible] \frac{D_e^2}{[illegible]} \frac{[illegible]}{D_e^2} = [illegible]$

$\frac{[illegible]}{[illegible]} = [illegible] \frac{V^2}{[illegible]} = \frac{[illegible]}{D^4}$

[illegible] 75% [illegible] steam [illegible] cost $[illegible]

[illegible] $= \frac{[illegible]}{D^4} \times \frac{[illegible]}{60 \times 33{,}000} = \frac{[illegible]}{D^4}$

$\frac{\text{steam costs}}{\text{hr.}} = \frac{[illegible]}{D^4} \times \frac{[illegible]}{.65 \times .75} \times \frac{[illegible]}{350 \times 1000 \times 773} = \frac{[illegible]}{D^4}$

$\frac{[illegible]}{D^4} \times 24 \times 360 = \frac{\$1.97 \times 10^5}{D^4}$

D ft	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	136	34736	8750	$79	[illegible]
[illegible]	14200	5.45	545	14745	3250	316	$3566
3 ft	3530	42	4200	7730	1700	2440	$4140
4 ft	7550	13.3	1330	8880	1940	770	$2710
4.5 ft	10540	8.35	835	11375	2510	483	$2993

[illegible]

$V = \frac{[illegible]}{[illegible]}$ [illegible]

$\frac{[illegible]}{[illegible]} = \frac{3.25}{[illegible]}\left(\frac{450}{[illegible]}\right)^2 = \frac{[illegible]}{D^4}$

[illegible] $= \frac{[illegible]}{D^4}$ [illegible]

D ft	[illegible] $162.6\,D^3+900$	H.D $\frac{1520}{D^4}$	[illegible] Turbines	[illegible]	F. C.	[illegible]	[illegible]
4	11400	5.95	595	11995	2640	344	2984
3	5300	13.8	1880	7180	1580	1090	2670
[illegible]	34[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
3.5 ft	7900	10.1	1010	8910	1960	[illegible]	2546

[illegible] $V = \frac{235}{D^4}$ [illegible] $= 49$ ft

[illegible] $= \frac{[illegible]}{D^4}$ $H.D = \frac{[illegible]}{D^4}$ [illegible] $= \frac{[illegible]}{D^4}$

[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
3	$[illegible]	[illegible]	[illegible]	11240	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	7785	[illegible]	[illegible]	[illegible]
2	[illegible]	[illegible]	[illegible]	6720	[illegible]	[illegible]	[illegible]

Water Gas Shift Reactors

Based on the calculations and references given below the following data have been prepared for the water gas shift reactors.

Investment cost (including supports, foundations, and installation) and initial catalyst cost)	$ 100,000
Steam requirement (680 psia and 500°F.)	20,000 #/hr.
Materials costs (for replacement of the iron oxide catalyst)	$1600 /yr.
Number of operators per shift	1

Nine identical reactors will be built; eight being required to process the normal load of gas and the ninth being a spare. The design data for each reactor are

Shell:

Height	38 ft.
Diameter (inside)	5 ft.
Thickness	1 in.
Material	steel

Tubes:

Length	35 ft.
Diameter (inside)	4.026 in.
Thickness	0.237 in.
Material	schedule no. 40 std. steel pipe
Number per reactor	61

Catalyst:

Volume of catalyst	180 cu. ft.
Space velocity (volume basis)	500 /hr.

Composition:

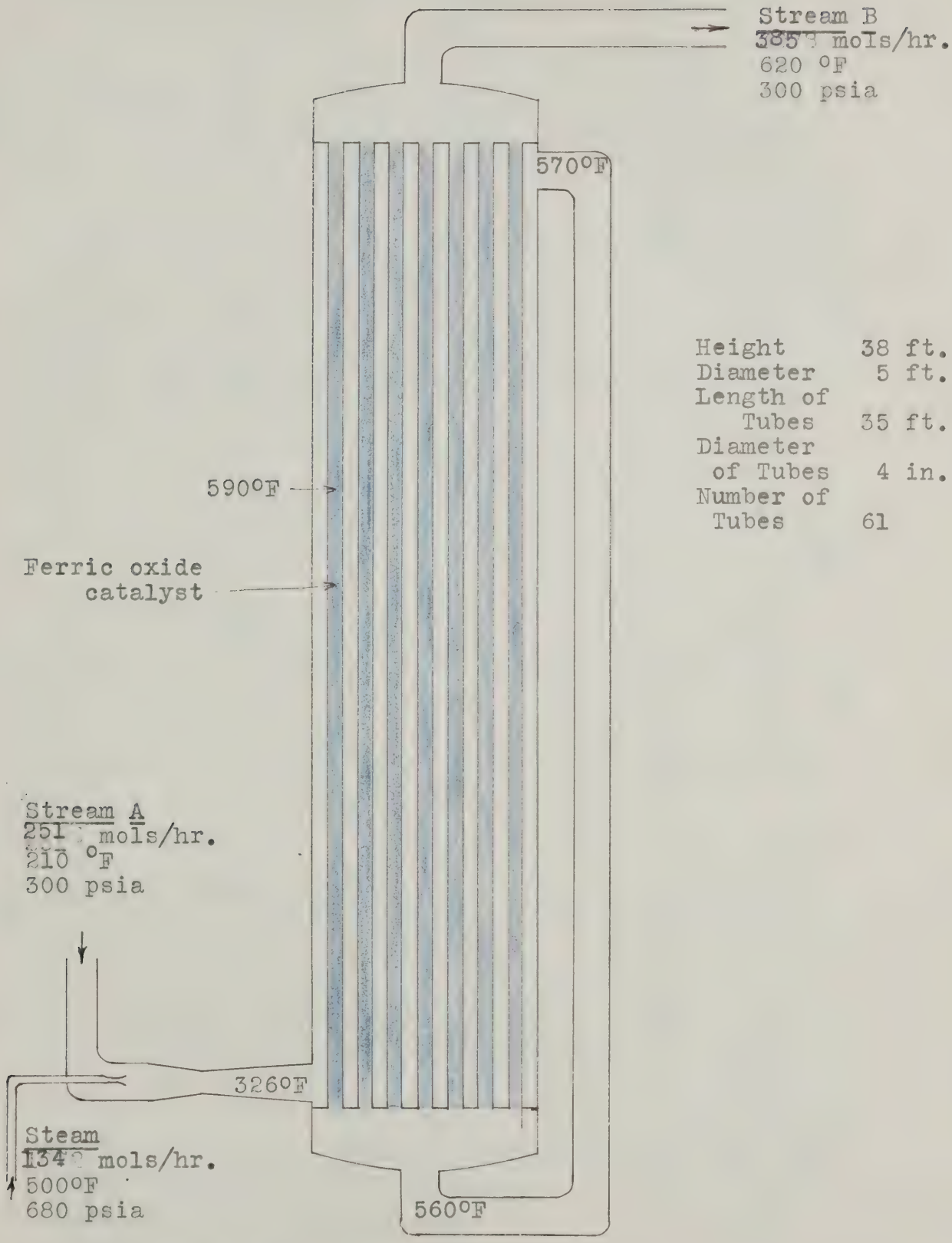
Compound	weight %
Fe_2O_3	54
MgO	17
$K_2Cr_2O_3$	7
Al_2O_3	
Na_2O	
SO_3	22
CO_2	
H_2O	

Size (avg. diam.) 0.125 in.

The reactors are of shell and tube design with the catalyst contained in the tubes and the gases being preheated in the shell side pass. The reaction steam is provided through a steam injector in order to mix it well and to recover the pressure available in it. It is assumed that the pressure rise in the injector pump will be large enough to balance the pressure losses through the reactor.

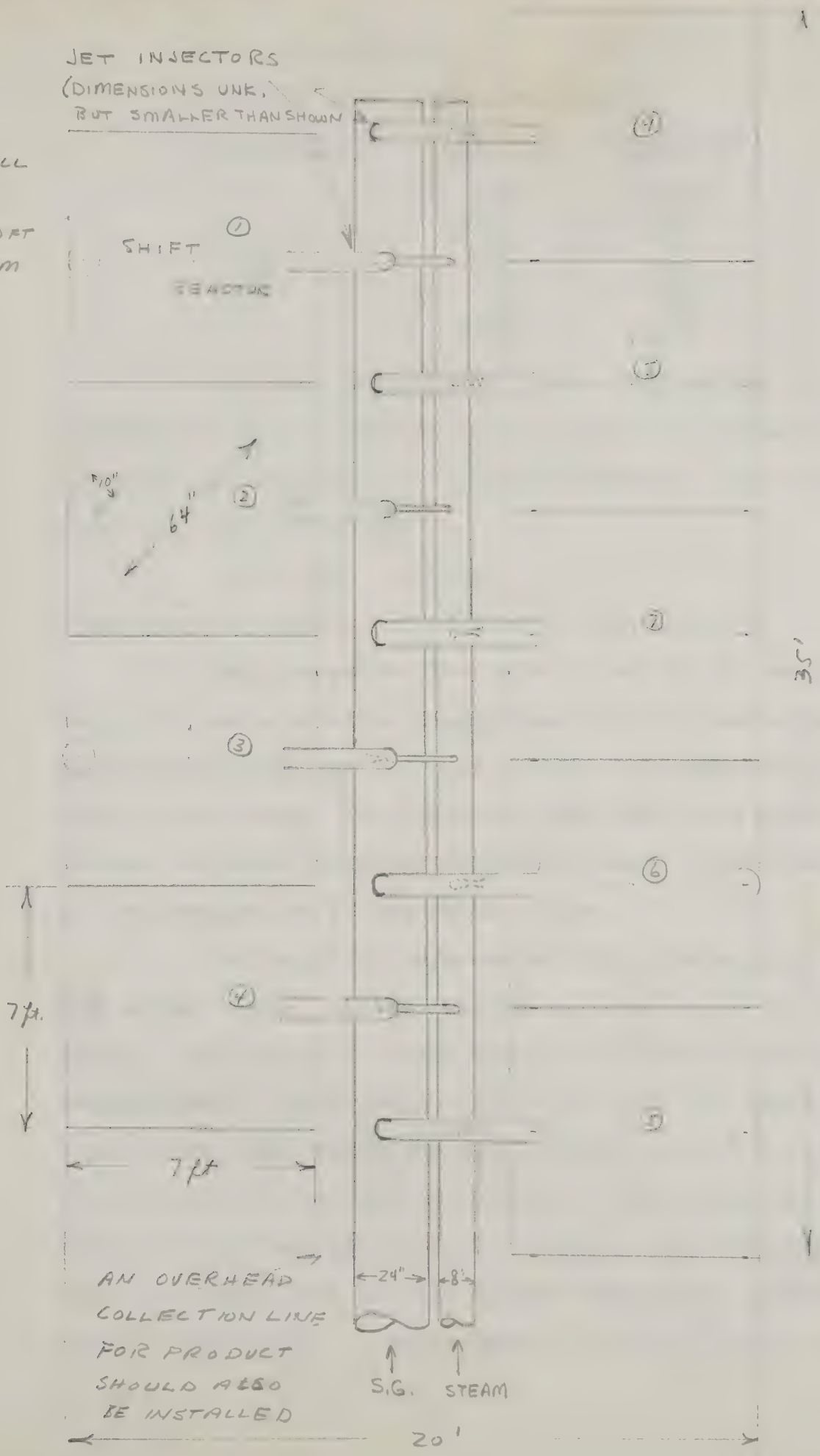
The following flow sheet contains all of the pertinent flow data. The stream compositions are as follows:

Component	volume %
Stream A	
CO	33.9
H_2	58.9
CO_2	0.9
H_2O	5.0
CH_4	0.1
N_2	1.2
Stream B	
CO	2.2
H_2	58.3
CO_2	20.5
H_2O	18.2
CH_4	0.1
N_2	0.7



Flow Sheet for One Shift Reactor

WATER GAS SHIFT REACTORS



Calculations

Incoming gas stream composition

Component	mols/hr.	volume %
CO	680	33.9
H_2	1182	58.9
CO_2	18	0.9
H_2O	100	5.0
CH_4	2	0.1
N_2	24	1.2
	2006	100.0

The requirement is to convert 90% of the CO entering to CO_2 by treatment with steam in the presence of a catalyst, which results in the production of H_2 by the water gas shift reaction

$$CO + H_2O = CO_2 + H_2$$

Therefore 612 mols/hr. of CO must be converted.

The gas enters the injector at 210 °F and is heated to the saturation temperature of the steam after expansion through the injector nozzle, by condensation of some of the steam. The mixed gas and steam then passes through the shell side of the reactor where it is heated to the temperature of the catalyst bed.

The reaction temperature will be determined by the amount of gas passing through the reactor in the following manner. The amount of steam added is dependent upon the temperature of the reactor, since the lower the reaction temperature, the higher the equilibrium constant for the reaction and the smaller the amount of steam required to obtain the desired conversion. However, the lower the reaction temperature is, the higher the heat of reaction which is exothermic. As designed, all of the heat released

by the reaction must be used by heating the gases and steam up to the reaction temperature. The amount of gas to be heated is a constant, while the amount of steam varies depending on the reaction temperature. The proper amount of steam is that which when mixed with the gases will require just as much heat as is released by the reaction. In order that this calculation be facilitated, the following distinct steps in the process were imagined:

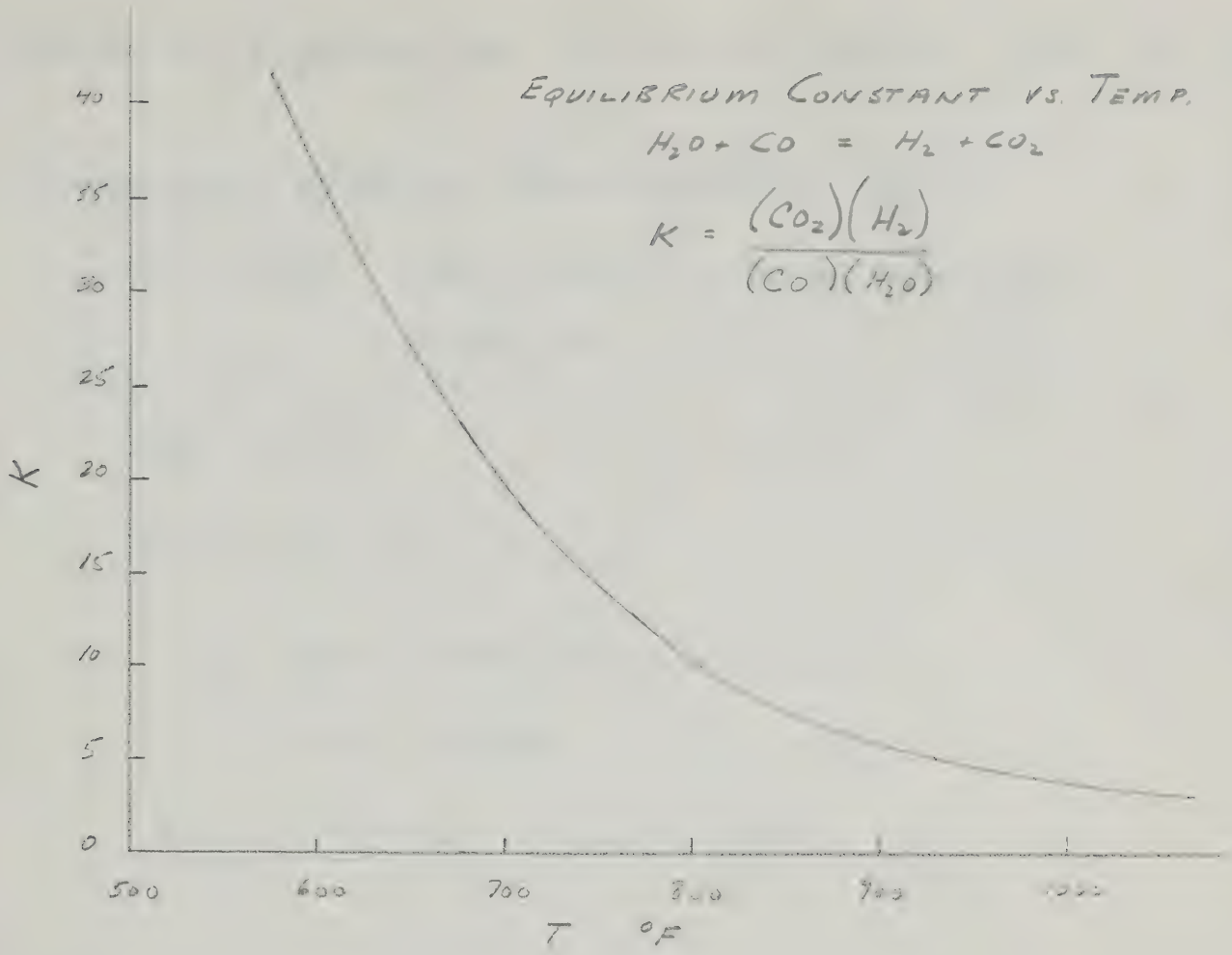
(1) Steam at 680 psia and 500 °F is expanded through the injector nozzle to the pressure of the incoming gas, 300psia. Eff. 85%

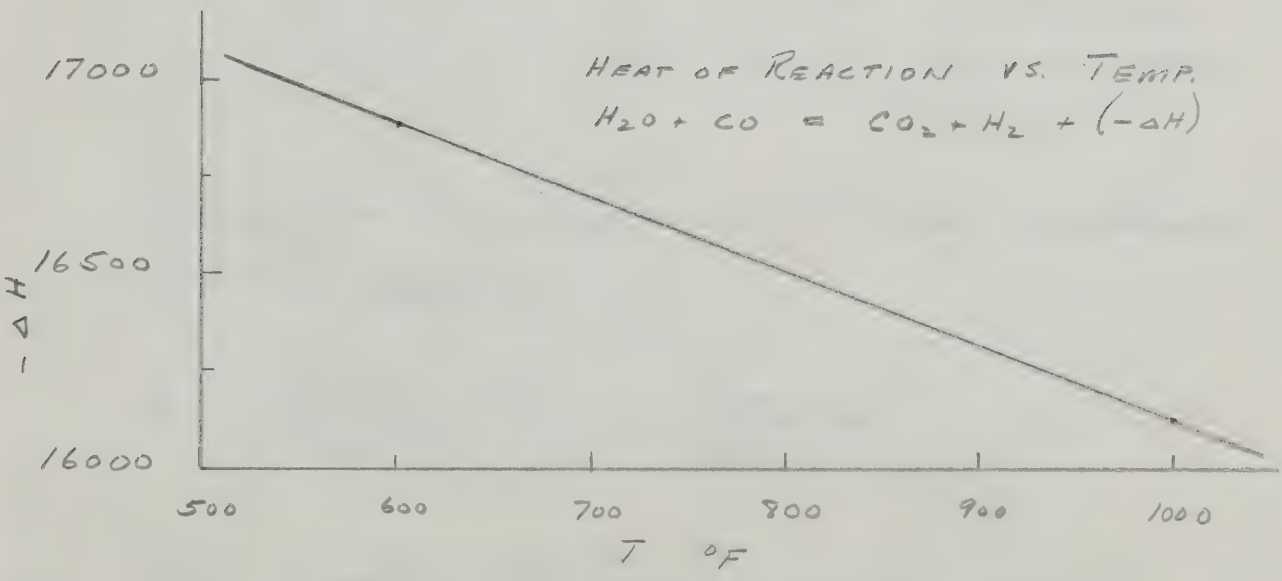
(2) Steam is expanded to its partial pressure in the steam and gas mixture with an efficiency of 60 %.

(3) Gas is heated from 210 °F to the saturation temperature of the steam after expansion by condensation of some steam.

(4) Steam and gas mixture is heated to the temperature of the catalyst bed by passing through the shell side of the reactor.

(5) Mixture reacts in contact with the catalyst to produce heat which is transferred to the incoming stream in step (4).





REF: CHEMISTRY OF COAL UTILIZATION p.p. 1589-1750

Steam is supplied at 500 °F & 680 psia i_1 = 1202 Btu/#

(1) Expansion of steam thru injector nozzle

$i_2 = 1135$ after constant entropy expansion to 300 psia.

eff = 85 %

$$i_2' = i_1 - e_1(i_1 - i_2)$$

$$= 1202 - 0.85(1202 - 1135)$$

$$= 1145 \text{ Btu/\#}$$

$x_2' = 92.8\%$ = quality of steam

<u>Trial 1</u>

Assume that reaction runs at 7000 F.

$$K = \frac{(CO_2)(H_2)}{(CO)(H_2O)} = \frac{(630)(1792)}{(68)(H_2O)}$$ if 100% of desired conversion were realized.

but in practice only about 98% of equilibrium will be realized

$$K = \frac{(643)(1792)}{(55)(H_2O)} = \frac{21000}{(H_2O)}$$

(H_2O) = mols of H_2O leaving reactor per hr.

$$(H_2O) = \frac{21000}{K_p}$$

at 700 °F $\quad K = 19.6$

$(H_2O) = 1070$

H_2O to be supplied = amt. leaving + amt reacted
- amt. entering

$= 1070 + 612 - 100$

$= 1582$ mols/hr.

$= 28500$ #/hr.

p_{S3} = partial pressure of steam in steam and gas mixture

$$p_{S3} = \frac{(.928)(1582)(300)}{(2006) + (.928)(1582)} = \frac{1468(300)}{3474} = 127 \text{ psia.}$$

$$i'_{S3} = i'_{S2} - 0.60\,(i'_{S2} - i_{S3})$$

$$= 1145 - 0.60\,(1145 - 1080)$$

$$= 1106 \text{ Btu/\#}$$

$t'_{S3} = 345$ °F

$\lambda'_{S3} = 874$ Btu/#

0.579 = specific heat of gas. in Btu/# x °F

Steam condensed to heat ~~to~~ gas to t'_{S3}.

$$\frac{(345-210)(0.579)(24700)}{(874)(18)} = 123 \text{ mols/hr cond.}$$

p_{S4} = partial pressure of steam after mixing and heat exchange.

$$= \frac{(1468-123)}{(3474-123)}\,300 = \frac{(1345)(300)}{3351} = 120 \text{ psi}$$

$$x_{S4} = \frac{1345}{1582} = 85\%$$

$$i_{S4} = 1059 \text{ Btu/\#}$$

Heat required to heat steam and gas to temperature of catalyst bed.

steam: $(28500)(1378-1059) = 9,100,000$ Btu/hr.

gas: $(24700)(0.579)(700-345) = 5,070,000$ Btu/hr.

$14,170,000$ Btu/hr.

Heat generated by reaction

$$(612)(16700) = 10,250,000 \text{ Btu/hr.}$$

New value of reaction temperature should be lower.

Trial 2

Assume that reaction runs at 600°F

$(H_2O) = \frac{21000}{36} = 583$ mols/hr H_2O leaving reactor

H_2O supplied $= 583 + 612 - 100 = 1095$ mols/hr.

$= 21900$ #/hr.

$$p_{33} = \frac{(0.928)(1095)(300)}{(2006) + (0.928)(1095)} = \frac{1017(300)}{3023}$$

$= 101$ psia

$i'_{33} = 1096$ Btu/#

$x'_{33} = \cancel{0.897}\ 89.7\%$

$$p_{33} = \frac{(0.897)(1095)(300)}{(2006) + (0.897)(1095)} = \frac{982(300)}{2988} = 98.5 \text{ psia}$$

$t'_{33} = 327°F$

$\lambda'_{33} = 889$ Btu/#

Steam condensed to heat gas to t'_{33}

$$\frac{(327 - 210)(0.579)(24700)}{(889)(18)} = 105 \text{ mols/hr.}$$

$$p_{S4} = \frac{(982-105)}{(2988-105)}\,300 = \frac{877\,(300)}{2883} = 93\ \text{psia}$$

$$x_{S4} = \frac{877}{1095} * 100 = 80\ \%$$

$$i_{S4} = 1009\ \text{Btu}/\#$$

Heat required to heat steam and gas mixture to reaction temperature

steam: $(21900)(1329-1009) = 7{,}000{,}000$ Btu/hr.

gas: $(24700)(0.579)(600-327) = 3{,}900{,}000$ Btu/hr.

$10{,}900{,}000$

Heat generated by reaction

$612\,(16880) = 10{,}320{,}000$ Btu/hr

A graphical plot of the results of trials 1 and 2 indicates an operating temperature of 585°F

Trial 3

Assume that reaction runs at 585°F.

$(H_2O) = \frac{21000}{39.2} = 535$ mols/hr H_2O leaving reactor

H_2O supplied $= 535 + 612 - 100$

$= 1047$ mols/hr.

$= 18820$ #/hr.

$$p_{s3} = \frac{(0.928)(1047)(300)}{(2006) + (0.928)(1047)} = \frac{970(300)}{2976}$$

$$= 98 \text{ psia}$$

$i'_{s3} = 1095$ Btu/#

$x'_{s3} = 89.6\%$

$$p'_{s3} = \frac{(0.896)(1047)(300)}{(2006) + (0.896)(1047)} = \frac{936(300)}{2976} = 95 \text{ psia}$$

$t'_{s3} = 324°F$

$\lambda'_{s3} = 892$ Btu/#

Steam condensed to heat gas to t'_{S3}

$$\frac{(324-210)(0.579)(24700)}{(892)(18)} = 102 \text{ mols/hr.}$$

$$p_{S4} = \frac{(936-102)(300)}{(2942-102)} = \frac{834(300)}{2840} = 87 \text{ psia}$$

$$x_{S4} = \frac{834}{1047} \times 100 = 79.7\%$$

$$i_{S4} = 1002 \text{ Btu/\#}$$

Heat required to heat steam and gas mixture to reaction temperature

steam: $(18820)(1323-1002) = 6{,}040{,}000$ Btu/hr.

gas: $(24700)(0.579)(585-324) = 3{,}730{,}000$ Btu/hr.

$9{,}770{,}000$ Btu/hr.

Heat generated by reaction

$$612(16940) = 10{,}360{,}000 \text{ Btu/hr.}$$

A graphical plot of the results of trials 2 and 3 indicates an operating temperature of 590°F

Trial 4

Assume that reaction runs at 590°F

$(H_2O) = \frac{21000}{37.6} = 560$ mols/hr. H_2O leaving reactor.

H_2O supplied $= 560 + 612 - 100$

$= 1072$ mols/hr.

$= 19320$ #/hr.

$$p'_{s3} = \frac{(0.928)(1072)(300)}{(2006) + (0.982)(1072)} = \frac{995(300)}{3001} = 99.2 \text{ psia}$$

~~$t'_{s3} = 326°F$~~

$i'_{s3} = 1097$ Btu/#

$x'_{s3} = 89.9\%$

$$p'_{s3} = \frac{(0.899)(1072)(300)}{(2006) + (0.899)(1072)} = \frac{964(300)}{3001} = 97 \text{ psia.}$$

$t'_{s3} = 326°F$

$\lambda'_{s3} = 890$ Btu./#

Steam condensed to heat gas to t'_{53}

$$\frac{(326-210)(0.579)(24700)}{(890)(18)} = 104 \text{ mols/hr.}$$

$$p_{54} = \frac{(964-104)(300)}{(2970-104)} = \frac{860(300)}{2866} = 90 \text{ psia}$$

$$x_{54} = \frac{860}{1072} = 80.1\%$$

$$i_{54} = 1010 \text{ Btu/hr.}$$

Heat required to heat steam and gas mixture to reaction temperature

steam: $19320(1325-1010) = 6,280,000$ Btu/#

gas: $24700(0.579)(590-326) = 3,770,000$ Btu/#

$10,050,000$ Btu/#

Heat generated by reaction

$612(16900) = 10,340,000$ Btu/#

590°F accepted as operating temperature

Reactor design

Space velocity = 500 ft^3 of water gas at std. cond. per ft^3 of catalyst per hr.

$$\frac{(2006)(359)}{500} = 1440 \ ft^3 \text{ of catalyst required.}$$

Length of tubes required to contain catalyst by sizes

Data is given for schedule no. 40 std. steel pipe

Nom. Size	I. D. (inches)	vol/ft (ft^3)	feet of tube req'd. to contain catalyst
4"	4.026	0.0882	16,300
3½	3.548	0.0687	21,000
3	3.068	0.0513	28,100
2½	2.469	0.0333	43,200
2	2.067	0.0233	61,800
1½	1.610	0.0141	102,000
1	1.049	0.0060	240,000

From pg. 174 McAdams "Heat Transmission" it is estimated that the empty tube heat transfer coefficient is

$$10 \ Btu/hr \times ft^2 \times {}^\circ F$$

from pg 178 McAdams "Heat Transmission" the multiplying factor for a packed tube appears to be

Hence the heat transfer coefficient is estimated as

$$60 \text{ Btu/hr} \times ft^2 \times {}^\circ F$$

for the inside surface of the tubes.

it is estimated that the heat transfer coefficient of the outside surface is

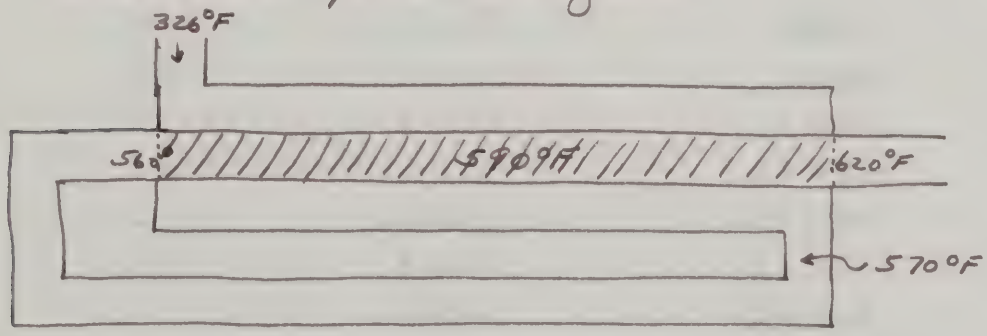
$$10 \text{ Btu/hr} \times ft^2 \times {}^\circ F$$

$$U = \frac{1}{\frac{1}{10} + \frac{1}{60}} = 8.58 \text{ Btu/hr} \times ft^2 \times {}^\circ F$$

Heat to be transferred

10,300,000 Btu/hr.

schematic flow diagram



$$\Delta t_{log\ mean} = \frac{(560-326)-(620-570)}{\ln \frac{560-326}{620-570}}$$

$$= \frac{234-50}{\ln \frac{234}{50}} = \frac{184}{1.543} = 119°F$$

Heat transfer area required

$$\frac{10,300,000}{(119)(8.58)} = 10080\ ft^2$$

Feet of tube of various sizes required to provide necessary heat transfer area.

Nom. Size inches	Outside area ft^2/ft	Feet of tube required
4	1.178	8560
$3\frac{1}{2}$	1.047	9640
3	0.917	11000
$2\frac{1}{2}$	0.753	13380
2	0.622	16200
$1\frac{1}{2}$	0.498	20200
1	0.344	29300

Costs of various size tubing, see p.497
Villbrandt, "Chemical Engineering Plant Design"

Nom. Size inches	cost/ft $	total cost of tubes to contain catalyst $
4	0.77	12,600
3½	0.67	14,100
3	0.57	16,100
2½	0.48	20,800
2	0.34	25,500
1½	0.25	25,500
1	0.16	38,400

From the above calculations it is clear that 4 in. std steel pipe should be used.

It appears that 35 ft. is longest practical length that can be used.

Therefore 465 tubes are required. These may be accommodated in 8 shells, 5 ft. inside diameter, being arranged with centers at the vertices of equilateral triangles 6½ inches on a side.

Hence 9 reactors will be built.

Actual heat transfer area per reactor

$$(61)(35)(1.178) = 2530 \text{ ft}^2$$

cost of shell per ft² of heat transfer area =

$$(1.75)(1.2)(0.60) = \$1.26$$

see Bliss, H Trans. AIChE 37,781 (1941)

cost of shell per reactor	=	$3320
" " tubes " "	=	1650

70,800 ft³/hr of gas must pass through the downcomer.

using nom. size 10" schedule no. 80 std. steel pipe will give a velocity of 40 ft/sec.

downcomer is 40 ft long and costs $6.00 per ft.

cost of downcomer = $240

Insulation:

max. allowable heat loss = 300,000 Btu/hr.

outside area of reactor with 6" of insulation

$$\frac{(40)(74)(3.14)}{(12)} = 775 \text{ ft}^2$$

max. allowable rate of heat loss

$$\frac{300,000}{775} = 387 \text{ Btu/hr} \times \text{ft}^2$$

thermal conductivity of 85% magnesia block = 0.05 Btu/hr × ft² × $\frac{°F}{ft}$

thickness of insulation required, taking Δt as 500 °F.

$$\frac{(0.05)(500)(12)}{387} = 0.78 \text{ inches}$$

One inch of insulation will be used.

Actual heat loss = 260,000 Btu/hr.

from Vilbrandt "Chemical Engineering Plant Design" p. 506

cost of 85% magnesia per board ft. = \$1.50

Actual outside area of reactor = 650 ft²

cost of reactor insulation = (650)(1.50)

= \$975

cost of insulation for downcomer = \$52.

Injector pump.

see Vilbrandt p. 474-5.

$P = e\,p\,r\,t + x$

$e = 1.3$
$p = 2000$
$r = 0.2$
$t = 0.8$
$x = 0$

$P = \$420$

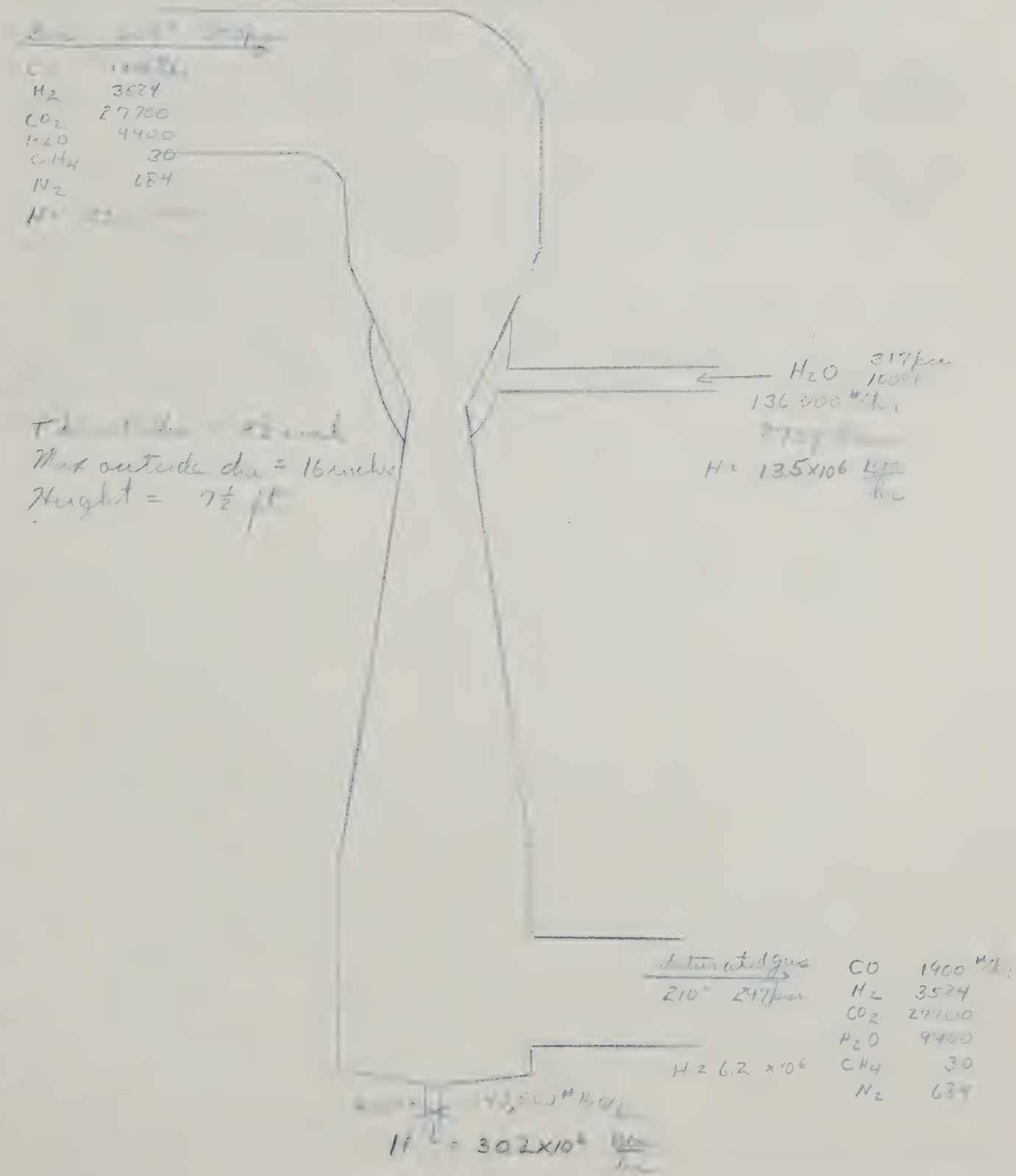
Total cost of one reactor:

	$
cost of shell	3320
" " tubes	1650
" " downcomer	300
" " insulation	1100
" " injector pump	420
	6790
+ about 50% for erection, supports, and foundations	3410
	$10200

Total cost of water gas shift reactor set

$9 \times 10200 = \$91800$

$\simeq \$100,000$



H2 3524
CO2 27700
H2O 9400
CH4 30
N2 684
Max outside dia = 16 inches
Height = 7½ ft
H2O 317 psi 100°F
136,000 #/hr
H = 13.5x10^6 Btu/hr
Saturated gas
210° 247 psi
CO 1900 #/hr
H2 3524
CO2 27700
H2O 9400
CH4 30
N2 684
H = 6.2 x10^6
H = 302x10^6 Btu/hr

Scrubber # 2 for gas from water [illegible] heater

The hydrogen [illegible] gas coming from the shift reactor is at [illegible] and then going to the [illegible] [illegible]. It should be cooled down at least to [illegible]. This scrubber dehumidifies the gas, reducing the water vapor pressure to [illegible] psi.

A discussion of the design factors for the [illegible] apparatus can be found in the [illegible] [illegible].

The final result gave a tower diameter of 2.2 [illegible], length of [illegible], [illegible] of [illegible] and [illegible] cost of [illegible].

[illegible] gallons of water per minute are needed at [illegible] and that water will be taken from the scrubber [illegible] [illegible] the gas [illegible] in which the water is heated up to [illegible].

Calculations

	In 600°F #/hr		Out 210°F #/hr	
[illegible]	68	1960	68	1960
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
	[illegible]	[illegible]	2646.3	[illegible]

Let water come in at 100°F & go out at [illegible]
v.p. of outlet gas will drop to [illegible]

Per hour this means 7560 # H_2O vap must condense out [illegible]

Basis 1 hr

[illegible]

$+ [illegible] + w(68) = 2340(1150) + (w + 7560)(1[illegible])$

[illegible]

[illegible]

[illegible]

[illegible]

[illegible]

[illegible]

[illegible]

Total Initial cost of scrubber =
(scaled down from scrubber #1)

Compressors for water to scrubber # 2

See discussion for compressors for water to scrubber # 1 for calculations and design factors. For this case the design for # 1 has not been scaled down from 1200 [illegible] [illegible] [illegible].

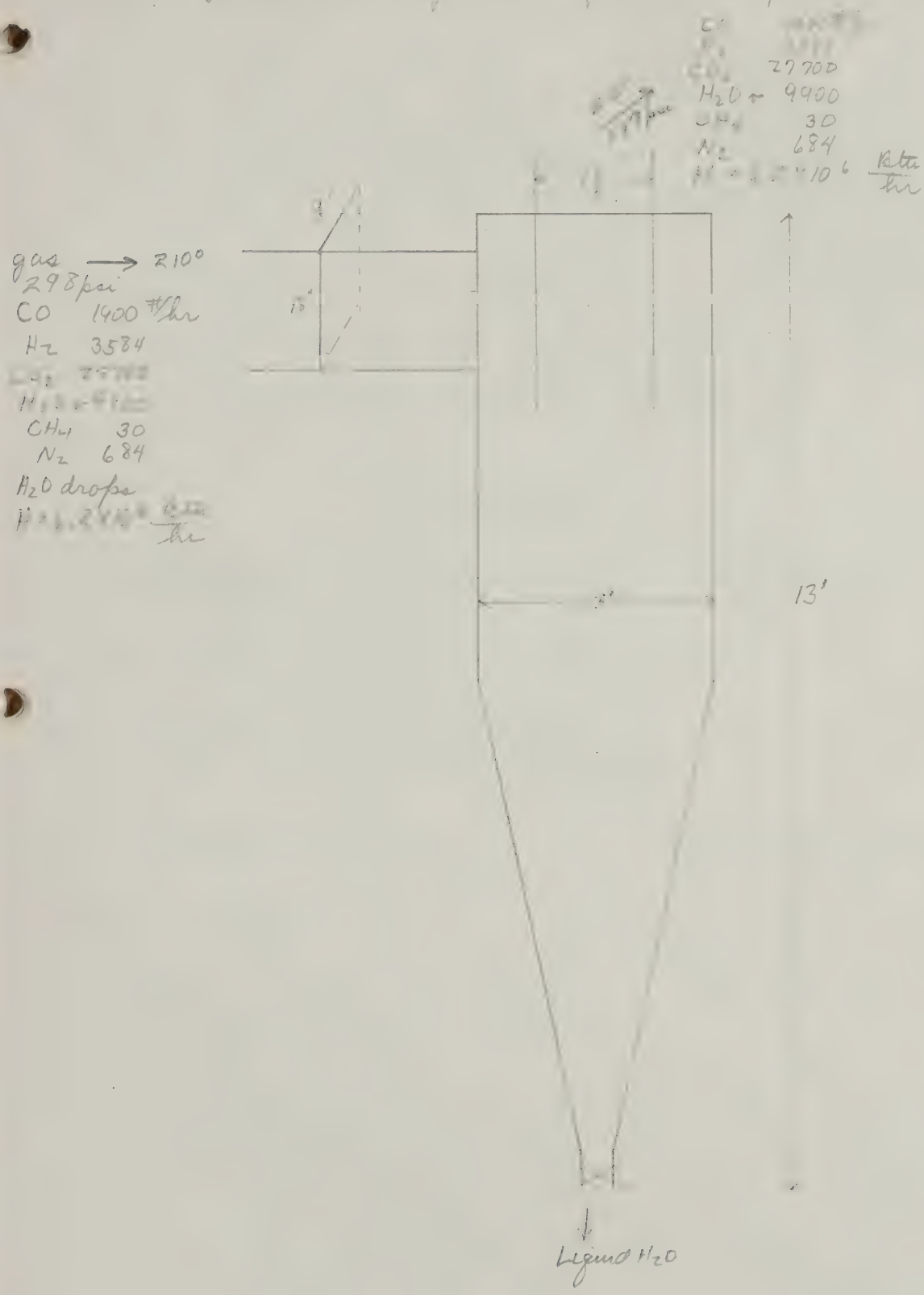
Stand by equipment will be [illegible]

Increasing rated horse [illegible] are
Pelton generator = 40 HP
Steam turbine = [illegible] HP
Ethyl compressor = [illegible] HP

Cost
Cost of 2-40 HP Pelton generators ($30/HP) = 80 X 30 = [illegible]
" " 2-30 HP Steam turbines ($100/HP) = 60 X 100 = [illegible]
" " 2-30 HP compressors ([illegible]) = [illegible] = [illegible]

Steam req'd = 265 #/hr

Cyclone #2 for Gas from Shift Reactor



Cyclone #2 [illegible]

Refer to discussion of Cyclone #1. In this case only 2646 $\frac{mol}{hr}$ [illegible] must be treated as [illegible] to 24,334 mol/hr [illegible]. [illegible] such that only 1 [illegible].

[illegible] to the [illegible] of the size & number of cyclones [illegible] the size of the cyclone [illegible] the [illegible] of [illegible] the [illegible].

[illegible]

[illegible]

$$1 - \frac{5}{6h} \quad \frac{[illegible]}{[illegible]} \quad \frac{[illegible]}{[illegible]}$$

$$\frac{[illegible]}{3} = \sqrt{\frac{[illegible]}{[illegible]}}$$

Dim = 3.9 ft [illegible]

Cost = $2000

Pressure drop = 2 psi

= CO_2 Absorption Tower Design =

CO_2 Absorption Tower

	mol
CO	63
H_2	1792
CO_2	19
H_2O	11
CH_4	2
N_2	[illegible]
	1897

gas 122°F
290 psia

amine solution with 0.2 $\frac{mol\ CO_2}{mol\ amine}$
122°F 300 psi
8265 mol soln / hr
25400 #/hr
1210 mol amine / hr
7055 " H_2O / hr

Dia of tower = 4 ft
total height = 90 ft

	mol
CO	[illegible]
H_2	1792
CO_2	[illegible]
H_2O	130
CH_4	2
N_2	24
	2646

200°
305 psi

305 psi

42% diethanol amine solution with 0.17 $\frac{mol\ CO_2}{mol\ amine}$

1210 mol amine / hr
7174 " H_2O / hr
348

CO_2 Bubble Plate Absorption Tower #1

[illegible]

[illegible]

[illegible]

[illegible] plant, and [illegible]

[illegible]

[illegible]

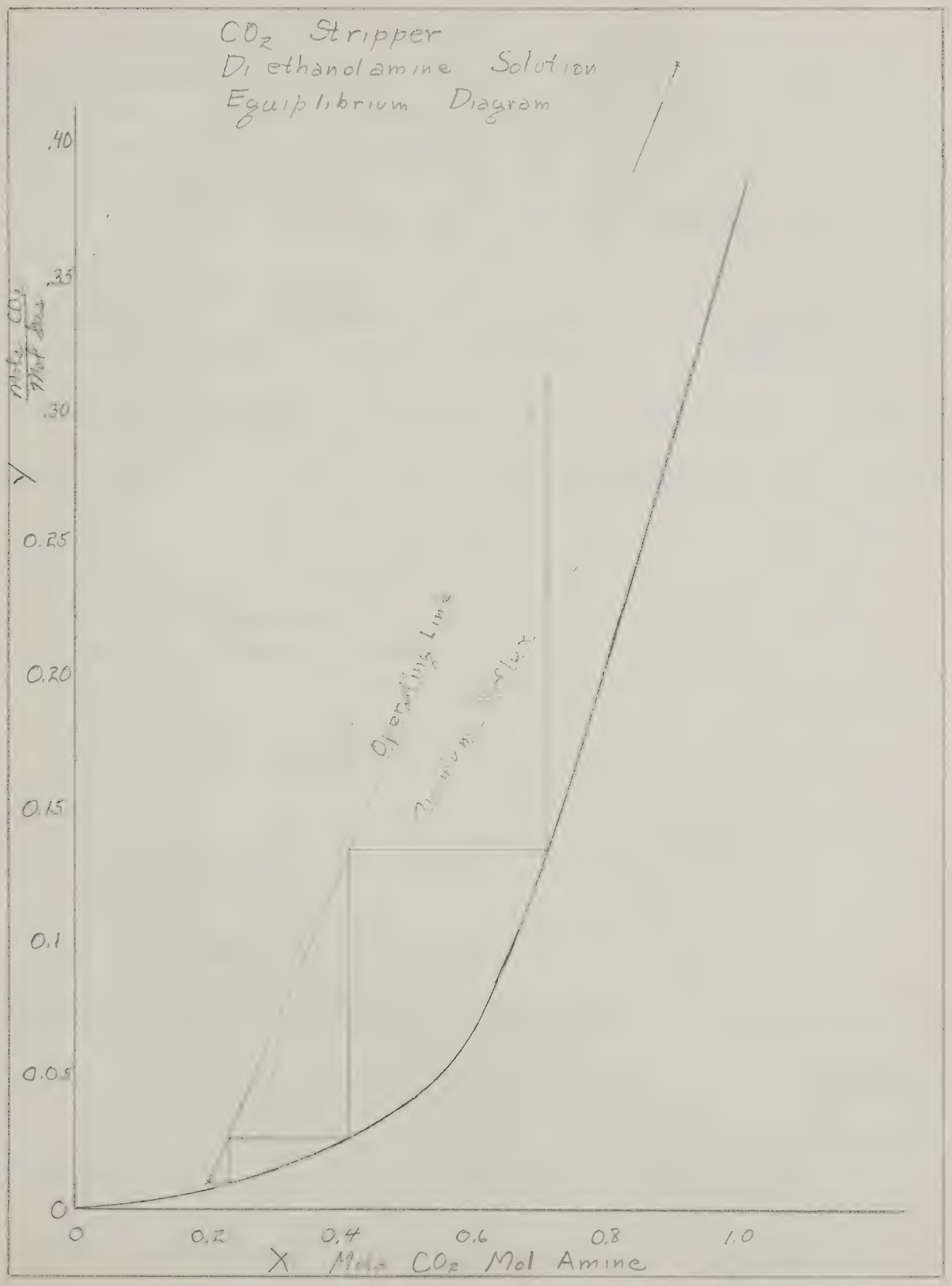
[illegible]

[illegible]

[illegible]

[illegible]

[illegible]


CO_2 Stripper
Diethanolamine Solution
Equilibrium Diagram
Y
.40
.35
.30
0.25
0.20
0.15
0.1
0.05
0
0
0.2
0.4
0.6
0.8
1.0
X Mol CO_2 Mol Amine

Calculations

For absorb 630-14 [illegible] 611 mol CO_2 [illegible] inlet amine solution containing [illegible] mol CO_2 [illegible] and outlet stream contains [illegible] mol CO_2 [illegible] equals $\frac{611}{[illegible]}$ = [illegible] mol amine/hr
= 8265 [illegible] amine solution/hr
= [illegible] #/hr

Heat balance - [illegible] assume [illegible] for amine [illegible] $\frac{Btu}{\#\,°F}$

[illegible] = 18 × 8324 H − 8265 (90) 18

H = [illegible] Outlet Sp Temp = 130°F

Calculation of tower area
Recover [illegible] spacing factor H = [illegible]
Bottom Pressure factor [illegible]
ρ_g = [illegible] T = [illegible]

Q = [illegible] $\frac{cu\ ft\ gas\ [illegible]}{day}$

$A = \frac{[illegible]}{[illegible]} \sqrt{\frac{[illegible]}{[illegible]}} = \frac{[illegible]}{[illegible]} \sqrt{\frac{[illegible]}{[illegible]}}$ [illegible]

D = 3.85 ft

Thus 1 tower 4 ft in diameter will be used.

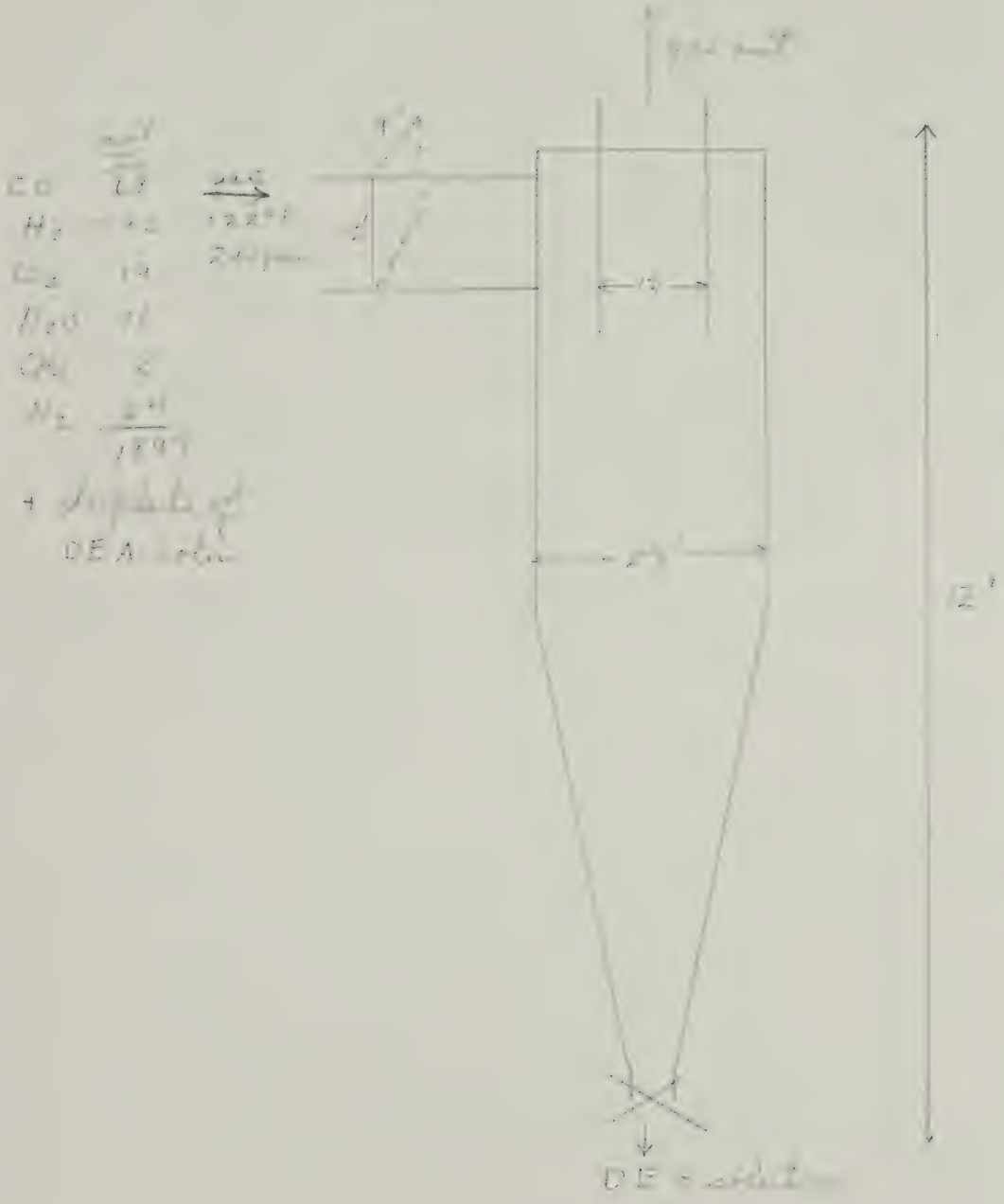
At [illegible] dollars/ft [illegible]

Cost of tower = [illegible]

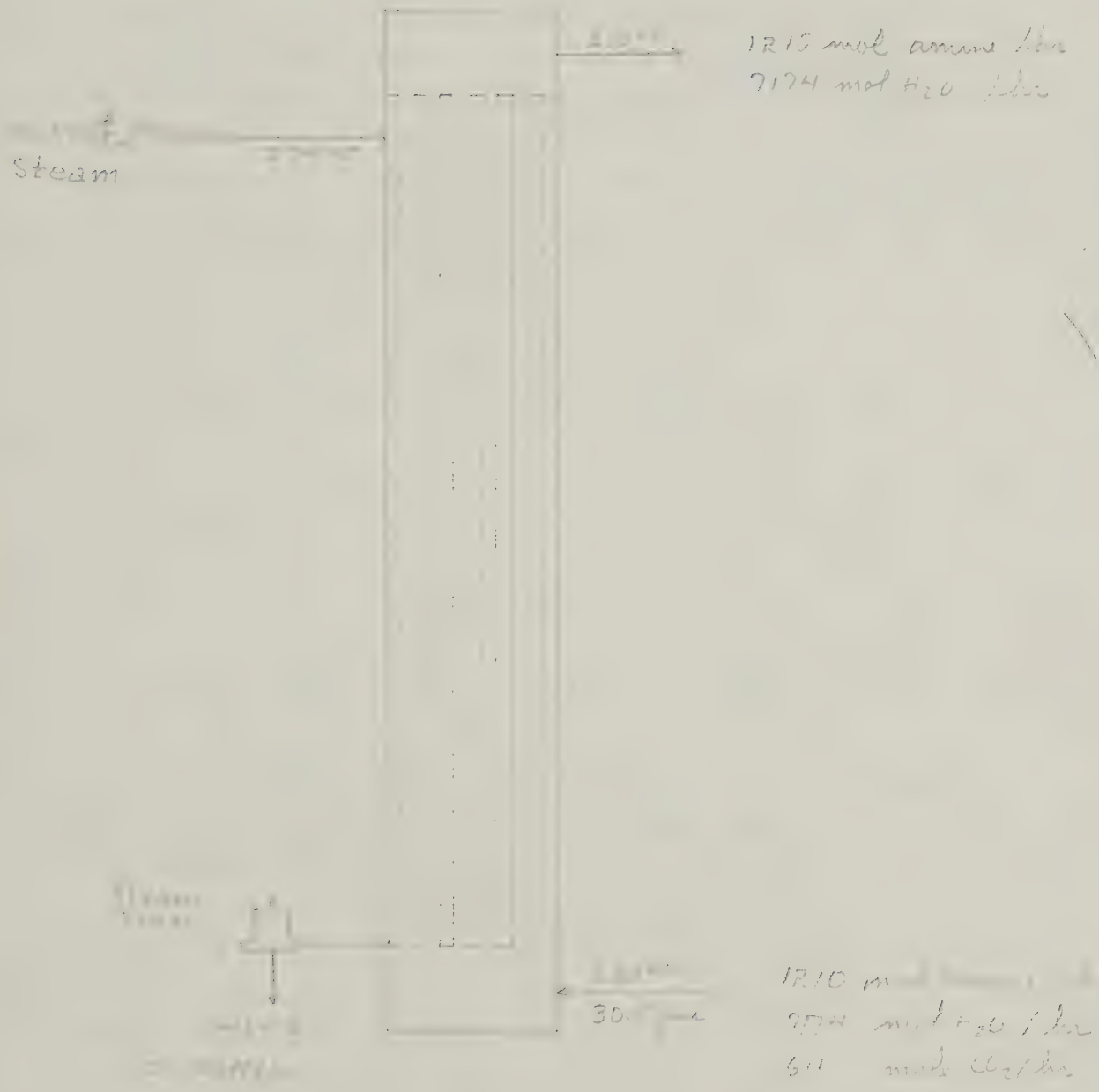
This includes support foundation [illegible]

Cyclones for Gas from CO_2 Absorption Column

2 Needed



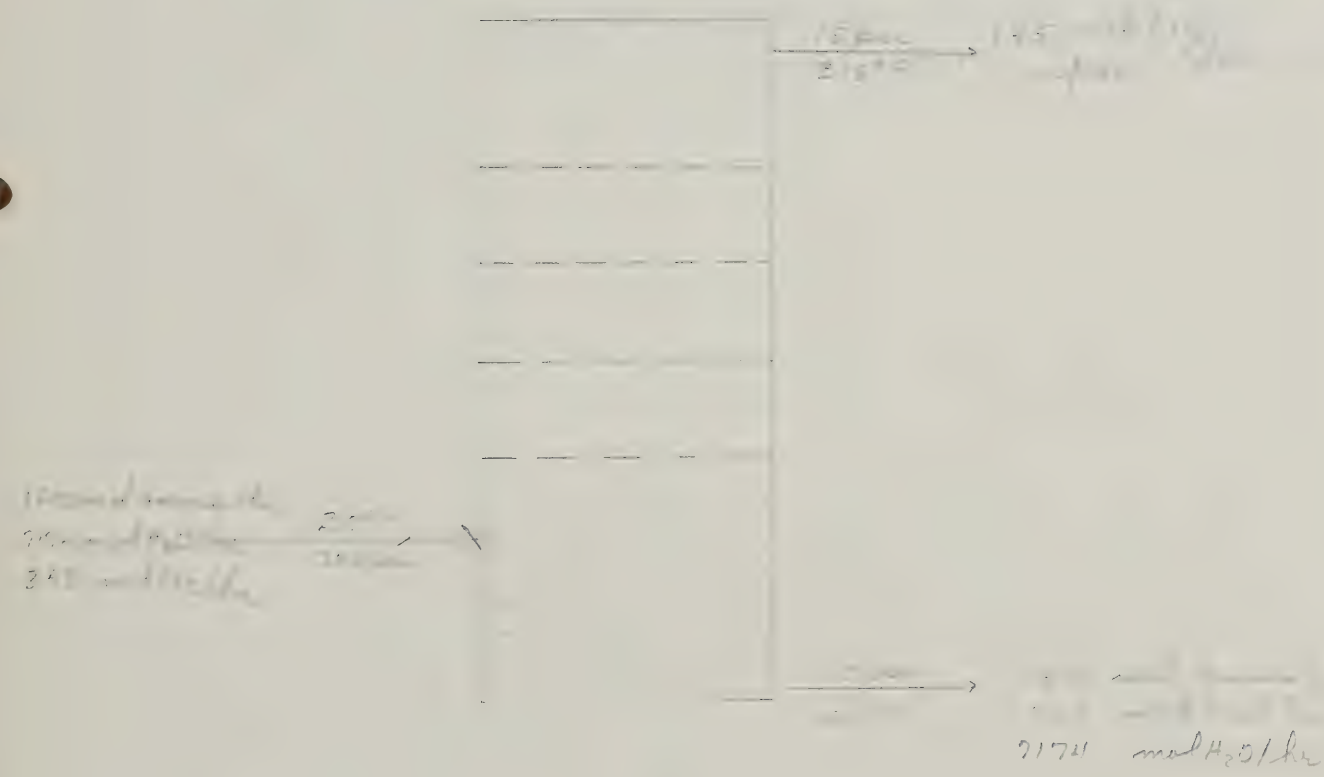
to Heat DEA Solution for Flash



in CO_2 [illegible] must be heated [illegible] the [illegible] tower.

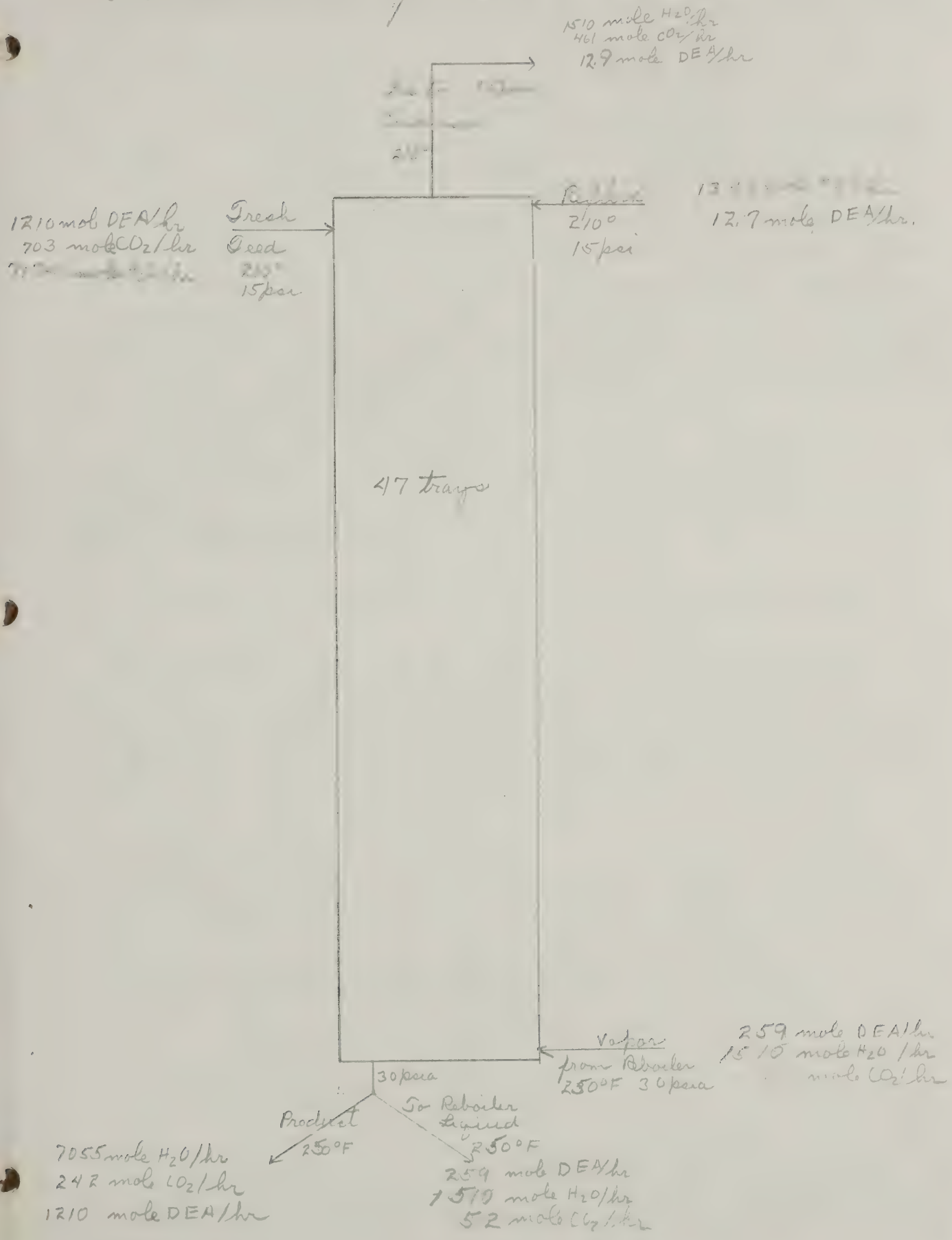
$$\frac{10,400,000}{[illegible]} = 8\%$$

Cooling water towers
35 P 6163 2
535.5
R53ze


21741 mol H_2O/hr

h_L = height of liquid in feet necessary

Distillation Tower for DEA Solution



[illegible] for DEA [illegible]

The liquid [illegible] from the [illegible] [illegible] with a ratio of [illegible] [illegible] of [illegible]. It is necessary in the [illegible] column to reduce this ratio to [illegible].

[illegible] part of X mole CO_2 per mole steam, versus X mole CO_2 per mole of amine in solution. [illegible] used a [illegible] L/G of [illegible] obtained. A [illegible] value of L/G = [illegible] [illegible] for the [illegible] value of [illegible] or [illegible] [illegible] of [illegible].

The [illegible] [illegible] [illegible] moles steam, and since the [illegible] [illegible] [illegible] moles of [illegible] [illegible] the top of the tower,

$$G = \frac{[illegible]}{[illegible]} = 1510 \text{ mol of steam/hr.}$$

The moles of DEA going out in vapor is based on a vapor pressure of the DEA of [illegible] mm. This is determined from a pure vapor pressure of DEA of 34 mm and Raoult's law holding.

$$34 \times \frac{[illegible]}{[illegible]} = [illegible] \text{ mm}$$

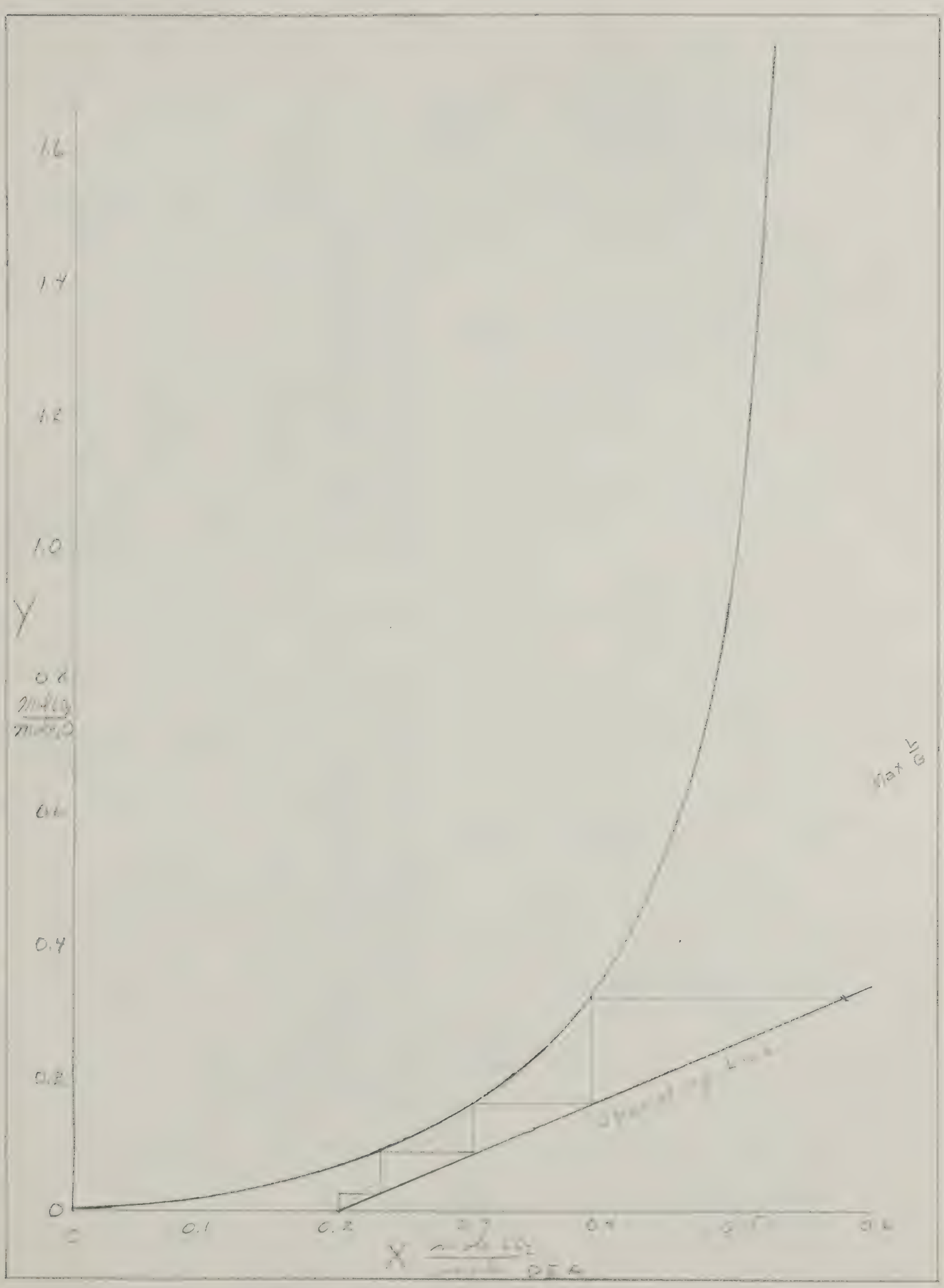
$$\frac{(1510 + [illegible]) \times [illegible]}{[illegible]} = [illegible] \text{ mol DEA}$$

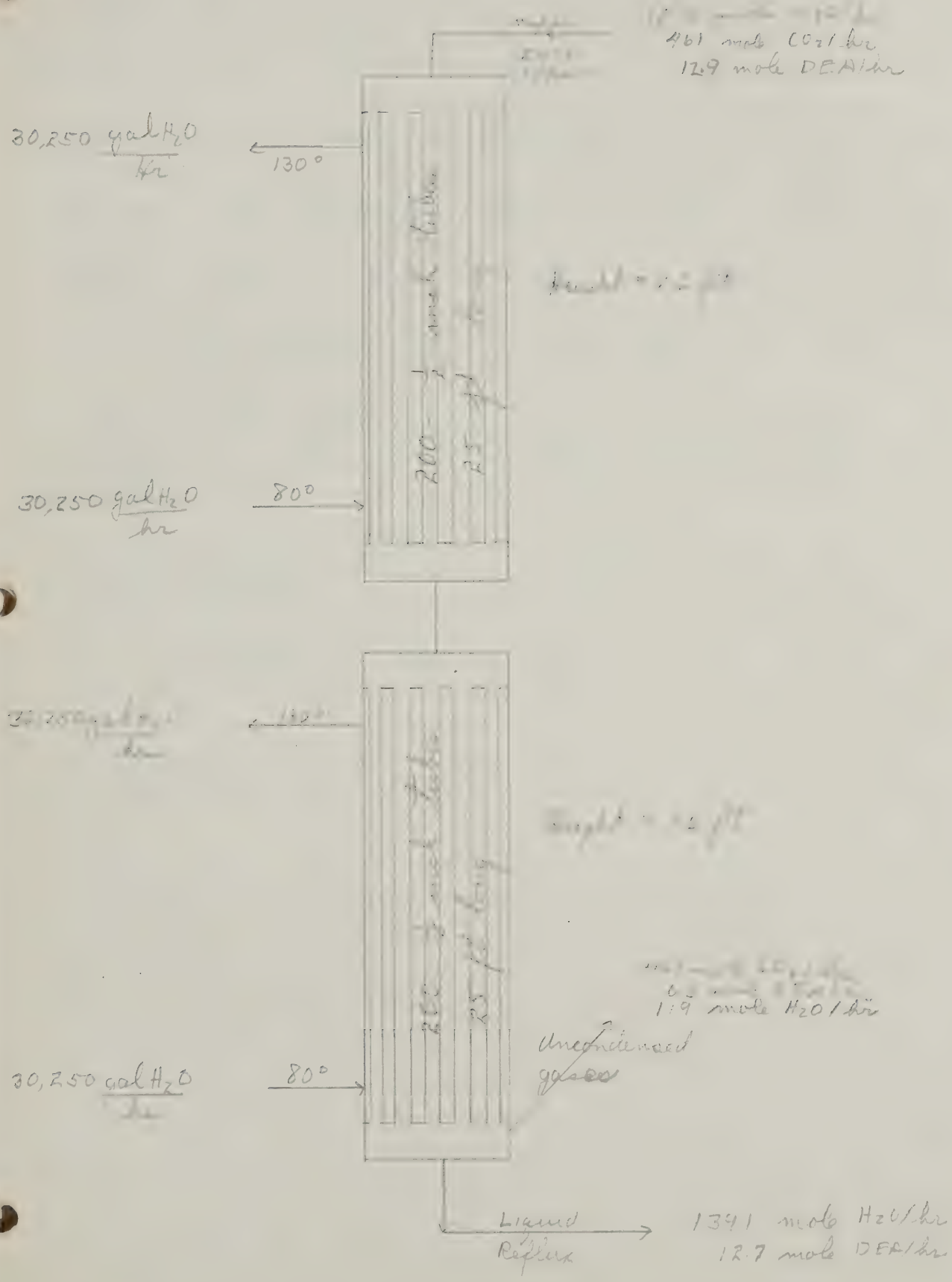
Since gas [illegible] the tower contains

1510 mole H_2O [illegible]
[illegible] " CO_2 [illegible]
[illegible] " DEA [illegible]

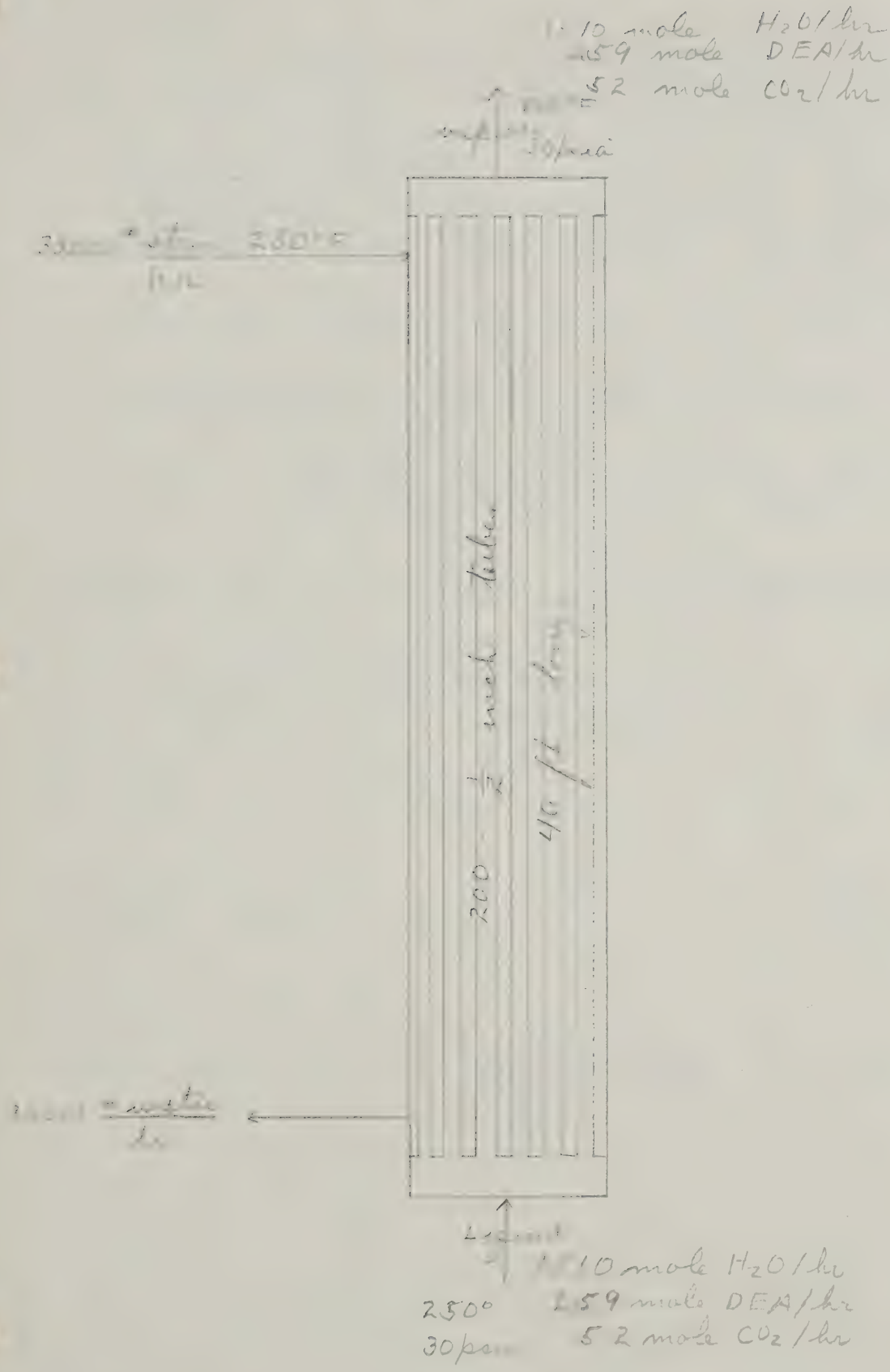
All but [illegible] mole H_2O should be condensed [illegible] [illegible] mol were condensed out of gas [illegible] [illegible] column.

1510 − [illegible] = [illegible] mol [illegible]


1.6
1.4
1.2
1.0
Y
0.8
0.6
0.4
0.2
0
0.1
0.2
0.4
0.6
Max L/G
Operating Line


461 mole CO2/hr
12.9 mole DEA/hr
30,250 gal H2O/hr
130°
200 ½ inch tubes
25 ft long
30,250 gal H2O/hr
80°
30,250 gal H2O/hr
130°
200 ½ inch tubes
25 ft long
1.9 mole H2O/hr
Uncondensed gases
30,250 gal H2O/hr
80°
Liquid Reflux
1391 mole H2O/hr
12.7 mole DEA/hr

Reboiler H-C 2



should be

[illegible]

[illegible]

[illegible]

[illegible]

[illegible]

[illegible]

[illegible]

[illegible]

[illegible]

[illegible]

[illegible]

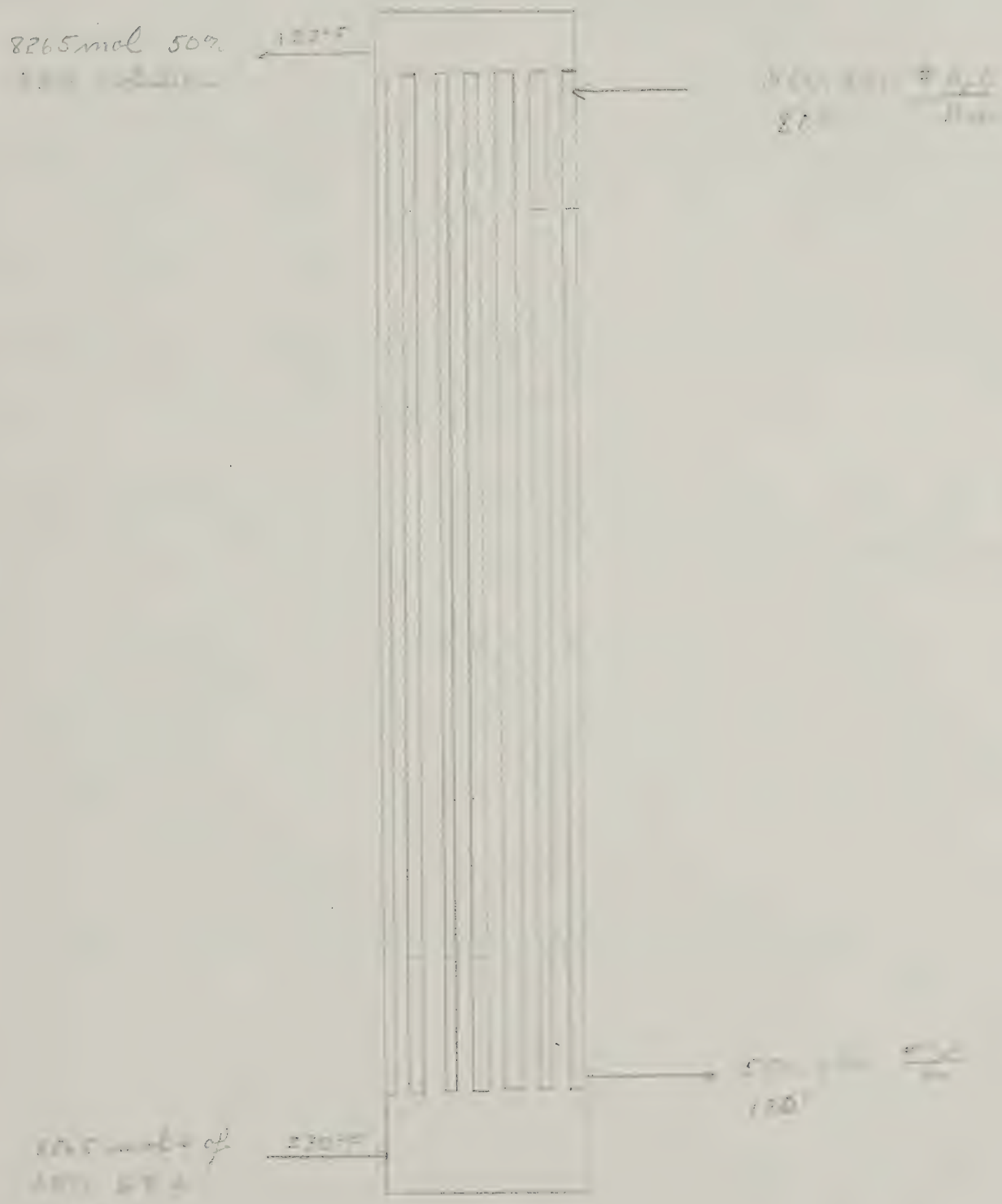
[illegible]

[illegible]

[illegible]

[illegible]

Cooler, H-C 3 for DEA to Absorber



Basic Equipment Cost of Synthesis Gas Plant
(Erection Labor and Freight Included)

Main Turbine and Compressors	$	2,300,000
H_2S removal unit		99,79[illegible]
Preheaters		348,000
Synthesis Gas Generators		371,500
First Venturi Scrubber unit		[illegible]
Water Shift Reactors		100,000
Second Venturi Scrubber Unit		[illegible]
CO_2 removal unit		214,320
Total	$	3,492,253

Steam Consumption

	650 p.s.i.g		50 psig	
Main Turbine and Compressors	165,000	lb/hr		
H_2S removal unit	1,900		5,055	lb/hr
First Venturi Scrubber unit	1,060			
Water Shift Reactors	20,000			
Second Venturi Scrubber unit	210			
CO_2 removal unit	1,700		46,400	
Total	189,870	lb/hr	51,855	lb/hr

Urbana, Illinois
May 31, 1950

Dr. H. F. Johnstone
Division of Chemical Engineering
University of Illinois
Urbana, Illinois

Dear Sir:

The following report includes all design specifications and costs for the section of a Fischer-Tropsch synthesis plant required for production of products from synthesis gas.

Any further information desired will gladly be furnished upon request.

Very truly yours,

Robert B. Field

Robert . Field
Reactor Design Group

REGENERATOR DESIGN AND CALCULATIONS FOR FISCHER-TROPSCH SYNTHESIS

Chemical Engineering Division
University of Illinois
Urbana, Illinois

Plant Design Group

R. Field
J. Hofto
D. McKinley
K. Timmerhaus

E. White

REACTOR AND REGENERATOR DESIGN

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REACTOR DESIGN

Summary.

It is hardly necessary to mention that the success of any chemical plant depends to a large extent upon the design and operation of the reactors. It is very probably due to this fact, however, that the literature on the products from the reactor, the design of the reactor, and the effects of the operating variables on the products is so limited and vague. For this reason the reactor design submitted here is only at best a reasoned estimate. The figures used have been selected only after as wide a search of the literature as possible. However, the necessity of picking up one fact here and another there has made any attempts at selection of an economic design almost impossible. The values used, however, are thought to be in almost every case somewhere near those being used or contemplated in actual plants.

The reactor system will consist of 4 fluidized reactors 11 ft. in diameter and 42 ft. high with 2.52 in. steel walls. They contain 2 in. chrome-manganese steam pipes to remove the heat liberated in the reactors. The reactors operate at 300 psia and 600° F. with a ratio of recycle to fresh feed of 3 to 1. 91,644 lbs./hr. of liquid product including hydrocarbons, water soluble, and oil soluble chemicals are produced from 23,230.5 mols/hr. of synthesis gas. The catalyst used will be iron mill scale reduced to an average particle size of 33 micron diameter. 136,800 lbs. of 680 psia steam are produced in the reactor per hour. Each reactor is fitted with a bank of multiclones to remove 95 per cent of the

5 micron diameter particles entrained in the product stream from the reactor.

The information on regeneration of Fischer-Tropsch catalysts is even more sketchy and vague than that on the reactors. In view of this, a regenerator has been designed which will remove carbon and wax from the catalyst by burning with air at atmospheric pressure and approximately 1200° F. The small amount of catalyst circulated per hour due to the small amount of fouling expected makes the recovery of any heat from the regenerator uneconomical. It must be emphasized, however, that the actual regeneration might be carried out in other ways or possibly would not be necessary or economical at all. Other methods mentioned are washing with oil and treating with hydrogen at elevated temperatures. It is felt that the regenerator as designed would be feasible for the latter method.

The total installed cost of the reactor system is estimated at $1,024,140. The operating cost exclusive of materials and compressor power is estimated at $158,114 per year, or 8.55¢ per barrel.

SUMMARY OF MATERIAL BALANCE

	#/hr.	mols/hr.
Fresh Feed	281,092	23,830
Recycle.	1,500,120	73,518
Regenerated + Fresh Catalyst . .	183,300	3,285
Water from Catalyst Reduction. .	8,000	500
Total to Reactors.	1,972,512	101,133
Effluent of Reactors	1,789,212	90,054
Catalyst to Regenerator.	183,300	3,285
	1,972,512	93,339

SUMMARY OF ENERGY BALANCE

All enthalpies referred to compounds in natural state at 32° F. and 1 atmosphere.

	Btu/hr.
Fresh Feed	45,800,000
Recycle Stream	32,550,000
Regenerated Catalyst Stream.	40,979,000
Heat of Reaction	681,000,000
Total Enthalpy Entering or Produced in Reactors . . .	800,329,000
Enthalpy of Product Stream	628,000,000
Enthalpy of Catalyst to Regenerator.	15,800,000
Enthalpy of Steam from Reactors.	156,529,000
	800,929,000

SUMMARY OF COST

I.	Reactors	$ 425,200
II.	Regenerator.	$ 2,953
III.	Pipes.	$ 42,185
IV.	Catalyst	$ 10,000
V.	Compressors.	$ 3,820
VI.	Instruments.	$ 21,900
	TOTAL EQUIPMENT.	$ 512,070
VII.	Installation	$ 512,070
	INSTALLED EQUIPMENT.	$1,024,140
VIII.	Depreciation	$ 51,207/yr.
IX.	Taxes, Insurance, Etc.	$ 51,207/yr.
X.	Labor.	$ 52,500/yr.
XI.	Fresh Catalyst	$ 1,200/yr.
	TOTAL OPERATING (less materials and power)	$ 156,114/yr.

INTRODUCTION

Considerable interest has been shown recently in the possibility of developing a process for making gasoline from natural gas at a cost competitive with the cost of gasoline from crude oil. Such a process would make natural gas this country's first important source of fuel independent of petroleum. The need for a substitute oil source is continually growing as the withdrawals from our crude-oil reserves continue to exceed the rate of discrovery of new fields. World War II naturally aggravated the situation immensely, but it remains a problem of real concern during the post war period because of our annually increasing rate of consumption. Just how long it will be before we face a really serious crude-oil shortage is a matter of conjecture, but the fact still remains: our oil reserves are dwindling.

The proven reserve of natural gas, has on the other hand, increased materially in recent years and there are indications that this trend will continue for some time. In fact, the present production rate is only half the discovery rate. If the production rate were doubled, according to Keith (1), approximating the rate of discovery, and if this increased production were converted to gasoline and oil, the motor-fuel production would be stepped up 25 per cent. Therefore, with a satisfactory natural gas-to-gasoline process like the Fischer-Tropsch process, it is evident that natural gas could become an important factor in the country's oil economy.

GENERAL DISCUSSION

History of the Process

In reviewing the history of synthetic gasoline, it is convenient to treat separately (1) the manufacture of the carbon monoxide and hydrogen mixture, and (2) the reaction of this mixture to produce gasoline and other hydrocarbons. This discussion will deal with the latter since the former will be amply covered by the gas synthesis group.

The catalytic reduction of carbon monoxide with hydrogen to produce hydrocarbons has been known since 1902 when Sabatier, the great French experimentalist, synthesized methane over Cobalt and Nickel catalysts at 400° - 600° F. In 1913, the Badische Anilin Soda-Fabrik disclosed that hydrogen and carbon monoxide would react at 100 - 200 atmospheres pressures in the presence of alkali-activated cobalt or osmium oxide supported on asbestos to form a liquid product containing alcohols, aldehydes, ketones, fatty acids, and some saturated and unsaturated aliphatic hydrocarbons. This work led eventually to the development of the methanol synthesis by the same company. In 1932, Franz Fischer and Hans Tropsch (2) working in the fuel research laboratories of the Kaiser Wilhelm Institute, found that alkalized iron turnings catalyzed the production of an oily liquid from hydrogen and carbon monoxide at 100 - 150 atmospheres pressure and 325° - 425° C. This product was called synthol, and contained alcohols, acids, aldehydes, ketones, esters, and hydrocarbons. Fischer and Tropsch found that, at lower pressures of about 7 atmospheres, the oil obtained was almost completely

paraffinic and contained a small fraction only of oxygenated organic compounds. At the lower pressure, however, only a very low yield of oil was obtained.

Working under the auspices of the Ruhrchemie A. G., Fischer and his associates (3) examined hundreds of catalysts and finally concluded that cobalt, nickel, iron, and ruthenium constituted the most active metals and that thoria and the oxides of other rare earths were good prometers. It was further shown that the presence of some alkali was desirable and that kieselguhr was the best support or carrier of the active ingredients.

By 1933, the process was considered to be well enough developed to build a commercial unit. Therefore, Ruhrchemie constructed a plant at Oberhausen - Holten with a rated capacity of 1,000 metric tons per year (equivalent to 8,000 42-gal. bbl.) of motor fuel and lubricating oil. The first catalyst used was a nickel-manganese-aluminum oxide mixture on kieselguhr; this was later supplanted by a cobalt-thoria-kieselguhr mixture. This plant, like others which followed it, had many inefficiencies which could be tolerated only under a structure supported by government subsidy. Among the early shortcomings of these plants especially with regards to the reactor were the following:

1. The catalyst was extremely sensitive and had to be broken in very slowly according to a rigorous schedule.
2. The initial cost of the catalyst was high and reworking was expensive.
3. The catalyst was rapidly poisoned by small traces of sulfur.
4. Catalyst life was limited to 6-8 weeks, due in part to

accumulation of high-melting-point waxes on the surface.

5. Conversion of carbon monoxide to useful products, was only about 40 per cent of the theoretical, despite the use of low space velocities (maximum of 100 volumes of carbon monoxide per hour per volume of catalyst).
6. The yield of oil per unit of catalyst per unit of time was low, due to the low space velocity, which in turn was limited by the capacity for removal of the heat of reaction.
7. The reactors were complicated, extended surface exchangers, with cooling surfaces no farther than 12-15 mm. from any part of the catalyst. Since the process was operated at atmospheric pressure, the reactors were necessarily large.
8. Recovery of heat of reaction as useful power was low.
9. The gasoline produced had an octane number of about 40, too low to permit its use as a motor fuel without additional treatment.

From a study of the German designs and operating data, it becomes clear that the only way to lower the cost of the process is to improve the efficiency. The Fischer-Tropsch process is fundamentally sound, but the German process was lacking in good engineering design. Improving the design involves finding a cheaper and more efficient means of heat transfer and a catalyst that would operate at a higher space velocity, the catalyst and the reactor design being inter-dependent.

Synthesis Mechanism and Carbon Formation

Formation of carbon on iron and iron-copper catalysts by the reaction $2CO = CO_2 + C$ has been studied by various investigators (4). The most significant result of this work (in so far as the Fischer-Tropsch synthesis is concerned) is the fact that neither an iron-free nor a copper-free carbon deposit was obtained. These facts indicate that carbon is deposited in the crystal lattice of the catalyst, and the inability to obtain a copper-free carbon deposit from tests with an iron-copper catalyst shows that iron carbonyl formation will not explain the results. It is very probable that carbon is formed from carbon monoxide by way of iron carbide as an intermediate. Carbidic carbon diffuses rapidly throughout the crystal lattice and subsequently decomposes to yield elemental carbon, thus disrupting the lattice structure.

Pichler and Merkel (5) at the Kaiser Wilhelm Institut für Kohlenforschung showed that extensive conversion of iron catalysts to the carbide Fe_2C results in a highly active and durable Fischer-Tropsch catalyst. High hydrogen to carbon monoxide ratios in the feed gas reduce the rate of carbon formation. Alkalized iron catalysts form elemental carbon more rapidly than those containing little or no alkali.

If one assumes that formation of elemental carbon occurs by way of some form of iron carbide as an intermediate, the fact that increasing the hydrogen to carbon monoxide ratio in the synthesis gas reduces the rate of carbon formation may be explained in either of two ways:

1. The hydrocarbon synthesis and the formation of elemental carbon are competing for the available supply of carbide.

Higher partial pressures of hydrogen are known to increase the rate of hydrocarbon synthesis, and therefore the rate of elemental carbon formation would be expected to decrease with increase of hydrogen to carbon monoxide ratio. This explanation assumes that bulk carbide (as distinguished from surface carbide) is a necessary intermediate in the Fischer-Tropsch synthesis. As shown by recent work of Emmett and his associates (5) using radioactive carbon as a tracer, this assumption is very probably incorrect.

2. Formation of carbon-to-carbon bonds in the lattice of the catalyst must occur to obtain elemental carbon from metal carbide. If the diffusion of carbidic carbon to nuclei where such carbon-to-carbon bonding has started is retarded by penetration of the lattice by atomic hydrogen, the rate of elemental carbon formation would decrease with increasing partial pressure of hydrogen. This explanation, based on rate of diffusion of carbidic carbon in the lattice, may serve also to account for the effect of alkali and other promoters or impurities. Thus the presence of alkali tends to preserve a structure similar to that of the spinels, and it is possible that carbidic carbon diffuses more readily through such a lattice than through that obtained when little or no alkali is present.

Overall Economics

An analysis of the Fischer-Tropsch overall economics seems to indicate: (8)

1. The plant should be located close to a large gas reserve. Approximately 11,000 cu. ft. of natural gas are required to produce 1 bbl. of liquid hydrocarbon product. Foreseeable gas reserves should at least equal 10 years' plant requirements.
2. The plant should be located on water to achieve lowest-cost transportation for both kinds of products -- hydrocarbons and chemical compounds. Location on the Gulf Coast, providing water transportation to the industries of the East Coast and to the upper Mississippi Valley would be quite advantageous.
3. The plant should be operated in conjunction with a petroleum refinery to assure constant availability of skilled technical personnel for general supervision, and for joint utilization of existing refining and distributing facilities.
4. Conservation regulations of various states will enforce gas-oil ratios. Continued satisfactory operation of a Fischer-Tropsch plant, requiring large volumes of natural gas, would necessitate the location near to a presently operating oil refinery to handle the large quantities of oil produced under these gas-oil ratios.
5. A complete Fischer-Tropsch plant to produce a large volume of liquid hydrocarbons, it is presently estimated, would have ~~an~~ a payout time ~~amortization~~ of approximately 8 years. A combined Fischer-Tropsch and chemical plant would have an ~~amortization~~ payout time of 3 years.

This acceleration in financial return is due to increased values of the chemical compounds over the values of the hydrocarbons.

6. Initial calculations have indicated that operating costs for the separation and the recovery of the chemical compounds are low, regardless of the assumed initial value of the by-product stream from the manufacture of the liquid hydrocarbons.
7. Even though the total volume of oxygenated compounds produced from any one plant constitutes a small percentage of the total postwar consumption of the chemical industry, the operation of many such plants by the petroleum industry might saturate the market for such chemical compounds and would adversely affect the profitability of such plants.

Evaluated under present economic circumstances, the factors of investment, of gas reserves, of cheap transportation, of marketing considerations, and of understanding the chemical industries, will create limitations which will undoubtedly serve to caution the widespread participation by many oil companies.

In developing a preliminary design for a plant utilising the Fischer-Tropsch process, the selection of a reactor and catalyst is the final step. The much lower cost of iron and its greater availability seems to indicate strongly its use. As all plants projected in this country have selected iron as a catalyst it will be used here. The type of reactor to be used will be considered next.

Catalyst and Reactor Selection

The catalyst temperature must be controlled within 25°F to maintain activity and product distribution, a minimum quantity of cooling surface must be provided therefore on the basis of a reasonable transfer rate. This in turn fixes the maximum space velocity allowable, assuming that the catalyst activity is not limiting. For example, the heat liberation for 90 per cent reaction of 1 cu. ft. of synthesis gas is about 55 B.T.U.; at a space velocity of 100 the hourly duty would be 5,500 B.T.U. per cubic foot of catalyst space. With a transfer rate of 10 B.T.U./ sq.ft./ hr./ °F. and a mean temperature difference of 10°F., the cooling surface required would be 55 sq. ft.

It is fairly obvious, then, why space velocity in such a reactor as the Germans designed is definitely limited. Raising it will increase the heat load, and therefore the temperature so that either product distribution will suffer (i.e., fixed gas production will increase) or the catalyst will coke up. Also, even under ideal conditions in such a system, localized hot spots may still occur. In fact, there is experimental evidence that the time catalyst temperature in a fixed-bed system is at least 75° - 100°F higher than the observed gas temperatures and that at isolated points it is higher.

Since catalyst activity is seldom the limiting factor, the space-time yield of oil becomes potentially much greater than has ever been realized with more adequate heat removal. From a review of the American petroleum industry, especially the "catalytic crackers" one possible method of improving markedly the space-time yield of oil is

to use the "fluidized catalyst" technique. Catalyst, ground to a very fine powder (approximately 75 - 80 per cent finer than 325 mesh) is maintained in a state of turbulence by aeration with regulated quantities of gas. When the powder is properly prepared, it can be aerated or partially suspended by the gas and assumes many of the properties of a boiling liquid. For example, there is a constant motion of the particles, and while the surface of the bed of powder is reasonably fixed, local eruption of bubbles are evident. Furthermore, the powder flows freely, exerts a static head, and is compressible.

Among the advantages of the fluid catalyst technique are: (1) temperatures can be controlled within 5° - 10° F.; (2) the catalyst itself is utilized as a medium of heat transfer; and (3) processes using the fluid catalyst are continuous, since any decline in activity can be compensated by withdrawal of a portion of the catalyst and the replacement with fresh material. The fluid catalyst also allows regeneration of a fraction of the catalyst without any costly work stoppages. Thus the stream efficiency will be much higher than with the fixed bed catalyst.

However, the fluid catalyst technique also brings with it certain problems. The reactor designed for the fluidized catalyst must be able to withstand the abrasive action of the catalyst at the prevailing pressure and temperature. This entails a construction with high cost special steel. Also to provide for maximum fluidization and minimum pressure drop through the reactor careful attention must be given to the placement of the heat removal elements in the reactor. Obviously the surface area of the heat exchanger is fixed since the amount of

heat to be removed is a constant for a given throughput. However, an economic balance would have to be made to determine the optimum pipe size which will best fulfill the previously mentioned requirements.

The size of each reactor and thus in turn the total number of reactors is determined from economic considerations. Factors of pressure, temperature, additional supporting structure, diameter, height, and thickness must all be considered in the final cost evaluation. The optimum reactor volume under these considerations has been found to be approximately 3500 cu. ft.

The heat removal from each reactor presents numerous other problems besides the optimum size of piping. High pressure steam is generated in the heat exchanger. Considerable economics is involved to determine the optimum pressure and quality of steam which should be obtained from this heat exchanger. The requirements of various turbines, compressors, and other steam operating devices must be established and the steam output of the heat exchanger fitted accordingly. Obviously if high pressure steam is obtained the Δt will be less and consequently the overall heat transfer will be less. On the other hand, the turbines cannot be operated efficiently, if at all, with low pressure steam.

Consideration must also be given to the heat removal from the reactor other than by the heat exchanger. The possibility that it is more economical to remove part of the heat from the catalyst regeneration stream or from either the recycle or fresh gas stream certainly exists. Only an economic balance here will give the answer. The overall heat transfer coefficient will be the same in the catalyst regeneration stream as it is in the reactor provided the pressure is kept constant. Otherwise

at a lower pressure, the coefficient is also lower. This also raises another question as to whether the regeneration unit should be operated at the same pressure as the reactor or at some other pressure. Both methods have certain advantages and disadvantages. Compression of the reducing gas used in the catalyst regenerator is an important phase to be considered since regeneration requires 20 volumes of reducing gas to 1 volume of catalyst.

The last major problem concerns itself with the catalyst. Only an estimate can be made upon the catalyst attrition. The design of the catalyst separator, however, as well as the catalyst makeup is affected by this estimate. To complete the economics on the catalyst an examination must be made to determine the optimum per cent of catalyst which is to be regenerated per unit time and the advisability of removing the waxes on the catalyst along with the carbon in the regenerator unit or removing them in a special dewaxing unit. Both coke and wax deposition affect the activity of the catalyst directly.

The use of various purities of liquid oxygen in the original preparation of the synthesis gas has a direct bearing on the amount of bleed in the recycle stream. A balance between the increased cost of higher purity oxygen and smaller volume of bleed and accordingly higher amounts of product is highly essential to obtain the lowest possible cost for liquid hydrocarbons by the Fischer-Tropsch process. Other diluents in the recycle stream must be considered also, but none affect the final outcome as much as the nitrogen impurity in the liquid oxygen.

The optimum temperature and pressure inside the reactor is more or less fixed by the catalyst used and the products desired. The pressure

and temperature of the recycle and fresh feeds is set by economic working conditions of the products separation group and the gas synthesis group respectively.

Discussion And Determination of Fischer-Tropsch Product Composition Using An Activated Iron Catalyst

The determination of the product composition from the Fischer-Tropsch process under design is the result of many comparisons in the literature to obtain the activated iron catalyst with the best qualities. Information on the iron catalyst is quite general, but an integration of results from the German organizations of Kaiser Wilhelm Institute, Lurgi, Brabag, I. G. Farbenindustrie, Ruhrchemie, and Rheinpreussen and the American organizations of Hydrocol, Kellog, Stanolind, and the Bureau of Mines has given the results shown below. The best catalyst reported was a doubly promoted synthetic ammonia catalyst, Fe_3O_4, Al_2O_3, K_2O (97, 2.5 and 0.5% respectively).

Information concerning a fluidized iron catalyst process similar to this one is given by Storch (7). For a temperature of 300-320°C and a pressure of 20 atmospheres the space time yield* of C_3 + is 181 bbl/1000 cu. ft./ day and the liquid product distribution is respectively 75% gasoline, 15% Diesel oil, 1% wax, and 9% alcohols and acids. These crude approximations were modified where more specific information was available and broken down into various components.

From the information on the results of the Lurgi catalyst the olefinic content of the oil and gasoline were established. The Lurgi recycle operation gives a low output of alcohols. Since this was not

* Weight of total product excluding methane, ethane, and ethylene, per volume of catalyst per unit time.

consistent with the fluidized iron catalyst product it was increased. The weight percents established from this catalyst agree very well with results published by the Bureau of Mines (8).

The Fischer-Tropsch product with this revision now has the following weight percent composition:

Table I Weight Per Cent

Fischer-Tropsch Product Composition

Product	Weight %	% Olefins
$C_1 + C_2$	16	--
$C_3 + C_4$	19	70
Gasoline	42	75
Diesel Oil	12	60
Wax	1	--
Oil and Water Soluble Chemicals	10	--

The oil and water soluble chemicals may be further classified into various alcohols, esters, ketones, aldehydes, etc.. Typical chemical yields from the hydrocarbon synthesis are given by Eliot, Goddin, and Pace (12). It should be stressed that these yields given below are restricted to synthesis conducted over a fluidized iron catalyst derived from mill scale at temperatures of 550° - 650°F and a pressure of 300 psia. (These are the operating conditions of the reactor used in this design). Distribution of individual water-soluble chemicals is shown, while oil-soluble chemicals are presented by classes.

Table I Typical Chemical Yields in Weight

Per Cent of Total Chemicals Produced

Water-Soluble Chemicals	Per Cent
Alcohols:	
Methanol	.2
Ethanol	26.8
Isopropanol	.5
n-Propanol	6.3
Butanols	2.9
Pentanols and higher	.9
Total alcohols	37.6%
Aldehydes:	
Acetaldehyde	4.3
Propionaldehyde	1.7
Butyraldehyde	1.5
Total aldehydes	7.5%
Ketones:	
Acetone	5.4
Methyl ethyl ketone	1.6
Methyl propul ketone	.6
Methyl butyl ketone	.2
Total ketones	7.8%
Acids	
Acetic	13.1
Propronic	3.4
Butyric	2.4
Valeric and higher	.6
Total acids	19.5%
Oil-Soluble Chemicals	
Alcohols	9.1
Aldehydes and ketones	9.3
Acids	9.2
Total oil-soluble chemicals	27.6%

It is apparent from this tabulation that the major proportion of the oxygenated compounds is in the low-molecular-weight range, the water-soluble chemicals amounting to about three times the volume of oil-soluble chemicals. Alcohols predominate among the water-soluble chemicals, followed by acids and smaller amounts of aldehydes and ketones. All classes are predominantly aliphatic. Only traces of unsaturates are found. The percentage of branching increases with increase in carbon number, and among the branched compounds the simple 2-methyl isomers predominate.

The authros (12) have plotted the yields of alcohols and acids as a function of carbon number. The most significant aspects of this plot are the smooth yield curve extending through the transition range (C_4 - C_5) from water-to-oil-soluble materials, and occurrence of maximum chemical yields in the two-carbon atom range.

With a knowledge of the weight percent of each component in the product, average molecular weights may be assigned to the median composition. In the case of the C_1 - C_2 group, the average was weighted on the methane side because of the predominance of methane in this group. Different molecular weights were assigned to the saturates and unsaturates of each group in order to simplify calculations for the product separation group.

The total number of mols of each constiuent in the product and the mol per cent by volume follows directly from above using a basis of 100 pounds of product. The mols of each constituent and the mols of H_2, C, and O_2 in each constituent in the total hourly product are

essential for a material balance around the reactor. The usage ratio of H_2 to CO is also established from the total number of mols of H_2 and C in the product. This ratio was found to be 1.18 and is used throughout the remainder of the report.

Tables 3 and 4 are both based on 100 pounds of product.

Table 3 Liquid Product Composition

Product	Wt %	Av. Mol. Wt.	Mol/100# Product
$C_1 - C_2$	16	18	0.889
$C_3 - C_4$			
C_nH_{2n-2}	5.7	51	0.112
C_nH_{2n}	13.3	49	0.272
Gasoline (C_5-C_{10})			
C_nH_{2n-2}	10.5	107	0.098
C_nH_{2n}	31.5	105	0.300
Diesel Oil (C_{10}-C_{18})			
C_nH_{2n-2}	4.8	205	0.023
C_nH_{2n}	7.2	203	0.036
Wax (C_{18} -)	1.0	282	0.00354
Water-Soluble Chemicals			
Alcohols ($C_nH_{2n-2}O$)	3.76	52	0.0723
Aldehydes ($C_nH_{2n}O$)	0.75	52	0.0144
Ketones ($C_nH_{2n}O$)	0.78	60	0.0130
Acids ($C_nH_{2n}O$)	1.95	64	0.0305
Oil-Soluble Chemicals	2.76	116	0.0238
Total number of moles			1.887

Table 3 Continued

Product	Mol C/100# Prod	Mol H_2/100# Prod	Mol O_2/100# Prod	Mol %
$C_1 - C_2$	1.000	1.890		47.2
$C_3 - C_4$				20.4
C_nH_{2n-2}	0.392	0.504		
C_nH_{2n}	0.960	0.960		
Gasoline (C_5-C_{10})				
C_nH_{2n-2}	0.735	0.833		
C_nH_{2n}	2.250	2.250		
Diesel Oil (C_{10}-C_{18})				3.1
C_nH_{2n-2}	0.334	0.351		
C_nH_{2n}	0.522	0.522		
Wax (C_{18} -)	0.071	0.071		0.0
Chemicals	0.443	0.510	0.1061	8.2
Total	6.707	7.897	0.1061	100.0

Usage ratio of H to C

$$\frac{7.897}{6.707} = 1.18$$

i.e.

H : C :: 1.18 : 1

Table 4 Water And Oil-Soluble Chemical Composition

Chemical Product	Mol C/100# Prod	Mol H_2/100# Prod	Mol O_2/100# Prod
Water-soluble chemicals			
Alcohols ($C_nH_{2n-2}O$)	0.1632	0.2250	0.0361
Aldehydes ($C_nH_{2n}O$)	0.0340	0.0390	0.0072
Ketones ($C_nH_{2n}O$)	0.0390	0.0390	0.0065
Acids ($C_nH_{2n}O$)	0.0640	0.0640	0.0325
Oil-soluble chemicals	0.1428	0.1428	0.0178
Total mols	0.4430	0.5098	0.1061

Preliminary Report From The Reactor Design Group Covering The Material Balance Around The Reactor

Feed Gas Composition:

The synthesis gas to be used in this Fischer-Tropsch plant supplied by the Synthesis Gas group has the following composition:

Synthesis Gas Constituents	Mol %	Mols/hr.
CO	32.05	7,630.0
H_2	60.81	14,500.0
CO_2	.83	197.3
H_2O	5.00	1,191.0
CH_4	.09	22.0
N_2	1.22	290.2
	100.00	23,830.5

The H_2 with synthesis gas has a ratio of 1.9:1 for the H_2 to CO ratio.

The total standard cu. ft./ day of the synthesis gas supplied is 205,000,000 cu. ft./ day or

8,541,700 cu. ft./hr synthesis gas supplied to F-T plant.

Synthesis Gas Constituents	Cu Ft/hr.	# / hr.
CO	2,732,000	213,500
H_2	5,200,000	29,000
CO_2	70,800	8,680
H_2O	427,000	21,430
CH_4	7,900	352
N_2	104,000	8,130
	8,541,700	281,092

Overall Balance:

Basis: 100 moles of fresh feed.

Overall balance of O_2, H_2, and C.

Mols H_2 in feed.

$60.81 + 5.00 + .18 = 65.99$

Mols O_2 in feed.

$16.02 + .83 + 2.50 = 19.36$

Mols C in feed.

$32.05 + .83 + .09 = 32.97$

<u>Assume:</u>

90% overall conversion based on CO charged of the 10% not converted 9% is lost as CO in the bleed and 1% is lost in the reactor as carbon on the catalyst.

The usage ratio of H_2:CO is 1.18 to 1 in the product. (This will be shown to be a valid assumption in a later discussion.)

The amount of O_2 in moles is .108 mols per 100 mols feed. (This will be shown also to be a valid assumption in a later discussion.)

No. of mols H_2 available for reaction in the reactor $65.99 - 5.0 = 60.99$

(These are the mols of H_2 in the product, bleed, and H_2O formed)

No. of moles C available for reaction in the reactor

$32.97 - (.10)(32.05) - .83 = 28.93$

(These are the mols of C in the product and CO_2 formed)

$(.09)(32.05) = 2.885$ mols C bled as CO

$(.01)(32.05) = .320$ mols C used for C deposition on catalyst

No. of moles of O_2 available for reaction in the reactor

$$19.36 - \underbrace{1.44 - 2.5 - .83 - .106}_{\text{this includes the } O_2 \text{ loss in the CO bleed, } O_2 \text{ in unreacted } H_2O, O_2 \text{ in unreacted } CO_2, \text{ and } O_2 \text{ in the product* (see further discussion)}} = 14.48$$

Reactor Reaction:

There are two reactions which predominate in the F-T reactor according to an intensive literature review by the reactor design group. These can be listed as,

$$CO + 2\,H_2 \rightleftharpoons (CH_2)_n + H_2O \qquad (1)$$

$$2CO + H_2 \rightleftharpoons (CH_2)_n + CO_2 \qquad (2)$$

These reactions along with other minor reactions are quite possible when using catalysts of either cobalt, iron, nickel, or ruthenium. Since this F-T process must compete with petroleum refinery processes it is essential to use the most efficient but also the most economical catalyst. For American production of hydrocarbons from the F-T process it has been found advisable to use the iron catalyst especially when designs call for fluidized catalytic units. The exact nature of the iron catalyst will not be discussed here.

First investigations with the iron catalyst were made by the German scientists working at the Kaiser Wilhelm Institut. First results showed that reaction (2) was favored by the catalyst by as much as ten to one. However, later investigations brought this figure down to 1:1 and even to 1:3 or 1:4. The latest information from the literature from both the Bureau of Mines (Gov't) publications and articles on the Carthage F-T plant being built in Brownsville, Texas report that the

iron catalyst has been developed now that it duplicates the reaction selection which the Germans reported for their Cobalt catalyst. Their information shows that reaction (1) was favored to the extent of ten times the reaction obtained in reaction (2). Since this is essentially going in the right direction for the products desired it has been assumed that the 10:1 ratio for the reactions is not unreasonable and will duplicate the American F-T process more closely than any other ratio. Ratios of 6:1 and 8:1 were investigated but both were discarded in favor of the 10:1 ratio because the former ratios incurred larger losses of CO and H_2 in the bleed with a correspondingly lower yield of hydrocarbon product.

Individual Balances:

Let for a basis of 100 mols of feed,

x = mols of H_2O formed in reactor

y = mols of CO_2 formed in reactor

z^1 = mols of $(CH_{2.36})$ formed in reactor

w = mols of H_2 in the bleed

Then for the O_2 balance,

$\frac{1}{2}x - y = 14.48$

$w - 1.18z^1 - x = 60.99$

$z - y = 28.93$

$x = 10y$ for the reaction ratio of equation (1) to equation (2) of 10:1 (catalyst selection)

Solving the Equations:

y = 2.41 mols CO_2 formed in reactor

z^1 = 26.55 mols ($CH_{2.36}$) formed in reactor

w = 5.71 mols H_2 in the bleed

x = 24.13 mols H_2O formed in the reactor

z = 26.66 mols total product*

*where z = mols of product including the oxygenated cmp'ds in the product.

Bleed Composition:

From the above balances it is now possible to give the values of the bleed stream. This computation assumes that all of the product is condensed out and the H_2O removed by A.P. Scheuler and his products separation group. That this cannot be done 100% it is fully realized and later allowances will have to be made for these traces which are not removed. The interest here is primarily to obtain reasonable figures which can be used in the direct design of the reactor and complimentary equipment.

Bleed Constituents	Mols/100 mols feed	Mol %
CO_2	3.24	24.83
N_2	1.22	9.35
H_2	5.71	43.72
CO_4	2.88	22.10
	13.05	100.00

It is noted that the recycle used here will have the same mol % composition as the bleed above.

Bleed Constituents	Mols/hr. Proposed	Std. ft^3/hr Proposed	#'s/hr Proposed
CO_2	773.5	277,500	34,000
N_2	290.7	104,400	8,150
H_2	1,361.0	489,000	2,722
CO	688.0	246,700	19,270
	3,113,2	1,117,600	64,142

Recycle Ratio:

A study of the Carthage F-T plant reveals that it is proposed to operate with a recycle ratio of 2.14 : 1. This indicates that the percent conversion per pass cannot be greater than 28.6. However, no additional data ~~is~~ are present which discuss~~es~~ the amount of catalyst degeneration which can be expected in the reaction. Thus in order to have a safety factor of nearly 50% it is proposed to operate at a recycle ratio of 3:1. This allows the reactor to produce the desired am't of hydrocarbons with a conversion per pass of 22.5%. It is believed that the increased costs due to a larger reactor will be completely balanced by the less stringent requirements which are placed on the catalyst regeneration unit, and the smaller heat exchanger unit needed to remove the heat of reaction inside the reactor.

Recycle Composition:

For a recycle ratio of 3:1 there will be 300 mols of recycle gas per 100 mols of feed gas.

Since the recycle has the same composition as the bleed gas, a direct calculation will give the mols of each constituent in the recycle stream.

Recycle Constituents	Mols/100 mols feed	Mol %
CO_2	74.49	24.83
N_2	28.05	9.35
H_2	131.16	43.72
CO	66.30	22.10
	300.00	100.00

Mols/hr. Proposed	St'd ft^3/hr Proposed	#'s/hr Proposed
17,770	6,370,000	781,500
6,690	2,400,000	187,200
31,224	11,210,000	62,450
15,807	5,670,000	442,600
71,491	25,650,000	1,473,750

Feed to Reactor Composition:

Reactor Feed Constituents	Mols/100 mols feed Recycle + Feed = Total	Mol %
CO_2	74.49 + .83 = 75.32	18.83
N_2	28.05 + 1.22 = 29.27	7.32
H_2	131.16 + 60.81 = 191.97	47.99
CO	66.30 + 32.05 = 98.35	24.59
CH_4	.09 = .09	.02
H_2O	5.00 = 5.00	1.25
	400.00	100.00

Mols/hr. Proposed	St'd ft^3/hr Proposed	#'s/hr Proposed
17,949	6,445,000	787,000
6,975	2,504,000	195,300
45,747	16,420,000	91,600
23,437	8,425,000	652,000
21	8,260	370
1,192	427,000	21,400
95,321	34,229,260	1,747,670

Reactor Product Composition:

Reactor Product Constituents	Mols/100 mols feed Feed + Reaction		= Product	Mols %
CO_2	75.32 +	2.41	= 77.73	21.18
CO	66.30 +	2.88	= 67.18	18.31
H_2	131.16 +	5.71	= 136.87	37.32
H_2O	5.00 +	24.13	= 29.13	7.94
N_2	29.27		= 29.27	7.98
C		.32	= .32	.001
Product		26.66	= 26.66	7.27
			367.17	100.00

Mols/hr. Proposed	St'd ft^3/hr Proposed	#'s/hr Proposed
18,520	6,650,000	817,000
16,000	5,740,000	449,000
32,620	11,710,000	65,291
6,975	2,495,000	126,400
6,990	2,504,000	195,300
76.3		915
6,360	2,282,000	91,664
87,541.3	31,381,000	1,745,570

Product Composition:

The total lbs. of product per 100 mols feed

$$(26.55)(14.36) + (.106)(32) = 384.65 \text{ \#'s product/100 mol feed}$$

Since the feed from the synthesis gas production group is 23,830.5 mols/hr, the total product in lbs. per hour will be

$$\left(\frac{23830.5}{100}\right)(384.65) = 91,664 \text{ \#/hr product.}$$

Constituents of Products	Wt. % of Products	Wt of Product #'s per 100 mols feed	Wt of Product #/hr Proposed
$C_1 - C_2$	16	61.50	14,660
$C_3 - C_4$ C_nH_{2n-2}	5.7	21.95	5,230
$C_3 - C_4$ C_nH_{2n}	13.3	51.22	12,210
Gasoline $C_5 - C_{10}$ C_nH_{2n-2}	10.5	40.35	9,620
Gasoline $C_5 - C_{10}$ C_nH_{2n}	31.5	121.20	28,880
Diesel $C_{10}-C_{18}$ C_nH_{2n-2}	4.8	18.45	4,395
Diesel $C_{10}-C_{18}$ C_nH_{2n}	7.2	27.67	6,590
Wax C_{18+}	1.0	3.85	917
Chemicals (H_2O Soluble)			
(a) Alcohols	3.76	14.46	3,445
(b) Aldehydes	.75	2.88	686
(c) Ketones	.78	3.00	715
(d) Acids	1.95	7.50	1,786
Chemicals (Oil Soluble)	2.76	10.62	2,530
	100.00	384.65	91,664

REACTOR SELECTION

A. GENERAL DISCUSSION

The feed to the reactor consists of one volume of fresh synthesis gas at 200° F and 325 psia, containing essentially hydrogen and carbon monoxide in a 1.9 to 1 ratio. This fresh synthesis gas is mixed with 3 volumes of recycle gas at 85°F and 325 psia. Regenerated catalyst at 1200°F is added to the reactor feed. This amounts to 183,300 lb./hr. of catalyst with enough hydrogen to prevent packing in the regenerated catalyst standpipe and to reduce any ferric oxide formed in the regenerator. The ~~combined~~ temperature of the combined streams is 171°F and the pressure is ~~171°~~ 325 psia.

There are four fluidized bed reactors each 11 ft. in diameter and 42 ft. high. On entering the reactors the feed is heated almost instantaneously to 600°F, the operating temperature of the reactors. The exothermic heat of reaction of the Fischer-Tropsch process makes necessary the removal of a large quantity of heat from each reactor to maintain the desired 600°F reactor

A. GENERAL DISCUSSION (CONT.)

Temperature. This heat is removed as heat of vaporization of 680 psia steam, generated in tubes located throughout the fluidized bed.

The product and unreacted gases are drawn off the top of the reactor at 600°F and 300 psia. The pressure drop through the bed necessary for fluidization being roughly 25 psia. Before leaving the reactor the product, in the gaseous state, passes through a bank of multicyclones to remove entrained catalyst. These multicyclones are designed to remove 95% of any 5 micron diameter particles present.

REACTOR HEAT BALANCE

In establishing a heat balance, the pure components in their natural state at 32°F and atmospheric pressure were used as the zero enthalpy state. The values of the specific heats and heats of vaporization of the various substances were determined by the products separation group. To insure uniformity in calculations these values were used whenever necessary. According to Storch ( ), the value of the heat of reaction for the Fischer-Tropsch Process is approximately 7,200 Btu./pound of liquid product. This value was taken to be the heat of reaction at 600°F, i.e. not including the heat of vaporization.

Enthalpy of Fresh Feed

Sensible Heat: $\Delta t = 168°F$

Constituent	Mols/hr.	C_p (AVG. 200-32°F) Btu/mol-°F	Btu/hr-°F	Enthalpy Btu/hr.
CO	7,630	7.21	55,000	9.24×10^6
H_2	14,500	6.92	98,900	16.60×10^6
CO_2	197	10.80	2,250	$.38 \times 10^6$
H_2O	1,191	8.60	10,230	1.72×10^6
CH_4	22	11.85	261	$.04 \times 10^6$
N_2	290.2	7.15	2,075	$.35 \times 10^6$

REACTOR HEAT BALANCE

Total Sensible Heat = 28.4×10^6 Btu/hr.

Only water is liquid at 32°F therefore:

Latent Heat = $14,600 \times 1,191 = 17.4 \times 10^6$ Btu/hr.

Enthalpy of Fresh Feed = 45.8×10^6 Btu./hr.

Enthalpy of Recycle Stream

Sensible Heat: $\Delta t = 53°F$

Constituent	Mols/hr.	Cp (AVG. 85-32°F) Btu/mol.-°F	B.t.u./hr.-°F	Enthalpy Btu/hr.
CO_2	17,750	10.8	191,000	10.12×10^6
N_2	6,690	7.15	47,800	2.53×10^6
H_2	31,260	6.92	195,000	10.33×10^6
CO	15,312	7.21	114,000	6.04×10^6
H_2O	144	8.60	1,240	$.07 \times 10^6$
CH_4	2,360	11.85	28,000	1.48×10^6
			577,040	30.55×10^6

Total Sensible Heat = 30.55×10^6 Btu/hr.

Only water is liquid at 32°F therefore:

Latent Heat = $144 \times 14,600 = 2.10 \times 10^6$ Btu/hr.

Enthalpy of Recycle Stream = 32.65×10^6 Btu/hr.

REACTOR HEAT BALANCE

Enthalpy of Regenerated Catalyst Stream

Sensible Heat : $\Delta t = 1168°F$

Constituent	Mols/hr.	C_p (AVG. 1168-32°F) Btu/mol-°F	Enthalpy Btu/hr.
Fe	3270	8.49	32,400,000
H_2	500	6.90	579,000
			32,979,000

There is a possible heat of reaction that may be liberated by the reduction of any Fe_3O_4 that may be formed in the regenerator. To allow for maximum heat input this will be considered.

$$Fe_3O_4 + 4H_2 \rightarrow 3Fe + 4H_2O + \Delta H_R$$

$$\Delta H_R = -64{,}000 \text{ Btu/H-mol.}$$

Heat of Reaction (maximum) $= -64{,}000 \times \frac{500}{4} = 9.00 \times 10^6$ Btu/hr.

Enthalpy of Regenerated Catalyst Stream = 40,979,000 Btu/hr.

Enthalpy of Catalyst from the Reactor to Regeneration

Any entrained material will be neglected

Sensible Heat : $\Delta t = 568°F$

Constituent	Mols/hr	C_p (AVG. 568-32°F) Btu/mol-°F	Enthalpy Btu/hr
Fe	3270	8.49	15,800,000

REACTOR HEAT BALANCE

Enthalpy of Product Stream:

Heat of Vaporization

Constituent	Mols/hr.	Heat of Vap. Btu/mol.	Heat of Vap. Btu/hr.
$C_3 - C_4$	1298	5,250	6.800×10^6
$C_5 - C_{10}$	1340	5,720	7.660×10^6
$C_{11} - C_{18}$	197	10,000	1.970×10^6
ALCOHOL	245	7,500	1.84×10^6
ALDEHYDE	49	7,500	$.368 \times 10^6$
KETONE	44	7,500	$.330 \times 10^6$
ACID	103	9,000	$.927 \times 10^6$
OIL SOLUBLE CHEM.	80	10,000	$.800 \times 10^6$
			20.695×10^6

91,664 lb. product/hr.

$$\Delta H_R = 7,200 + \frac{20,695,000}{91,664} = 7,430 \text{ Btu/lb.}$$

Constituent	Mols/hr.	C_p (AVG.) Btu/mol-°F	Heat of Vap Btu/mol.	Calc. of Enthalpy	Enthalpy Btu/hr
CO_2	18,520	11.1		$18,520 \times 11.1 \times 568$	116.900×10^6
CO	15,505	7.25		$15,505 \times 7.25 \times 568$	63.900×10^6
H_2	32,633	6.95		$32,633 \times 6.95 \times 568$	129.000×10^6
N_2	6,980	7.15		$6,980 \times 7.15 \times 568$	28.400×10^6
H_2O	7,618			$7,618 \times 1203 \times 18$	165.100×10^6
$C_1 - C_2$	3,000	13.20		$3,000 \times 13.2 \times 568$	22.500×10^6
$C_3 - C_4$	1,298	27.10	5250	$1298[5250 + 27.1 \times 568]$	26.800×10^6
$C_5 - C_{10}$	1,340	56.00	5720	$1340[5720 + 56 \times 568]$	59.200×10^6
$C_{11} - C_{18}$	197	105.	10,000	$197[10,000 + 105 \times 568]$	13.700×10^6
ROH	245	24.	7,500	$245[7,500 + 24 \times 568]$	5.120×10^6
RCHO	49	19	7,500	$49[7,500 + 19 \times 568]$	$.846 \times 10^6$

REACTOR HEAT BALANCE

Enthalpy of Product Stream (cont.)

Constituent	Moles/hr	C_p (AVG) Btu/mol-°F	Heat of Vap. Btu/mol.	Calc. of Enthalpy	Enthalpy Btu/hr.
RCOR	44	24	7,500	44[7,500 + 24X568]	0.929×10^6
RCOOH	103	21	9,000	103[9,000 + 21 X 568]	2.160×10^6
OIL SOL. CHEM.	80	43	10,000	80[10,000 + 43 X 568]	2.750×10^6

Enthalpy of Product Stream = 628.105×10^6 Btu/hr.

Heat Removed as Steam

Heat Liberated in reaction = 7,430 X 91,664 = 681.0×10^6 Btu/hr.

Enthalpy of Fresh Feed + Enthalpy of Recycle Stream + Enthalpy of Regenerated Catalyst + Enthalpy of Reaction = Enthalpy of Product + Enthalpy of Steam

Btu/hr.

$45.8 \times 10^6 + 32.5 \times 10^6 + 40.9 \times 10^6 + 681.0 \times 10^6 =$
$628.1 \times 10^6 + 15.8 \times 10^6 +$ Enthalpy of Steam

Enthalpy of Steam = 156.5×10^6 Btu/hr.

H steam at 500°F = 1201 Btu/lb.
H water at 500°F, 680 psia = 57 Btu/lb.
Enthalpy of Vap. / lb. = 1144 Btu.

Pounds of Steam / hr = $\frac{156.5 \times 10^6}{1144}$ = 136,800 # / hr.

Final Material Balance:

The Material Balance as presented in the Preliminary Balance does not take into account any components in the bleed except CO_2, H_2, N_2, and CO. The Preliminary Balance was prepared to give the Products Group something to work with in their design. They found it advisable to supply a recycle stream of the following composition:

Component	Mols/hr	St'd ft^3/hr.	#'s/hr.
CO_2	17,750	6,370,000	781,000
CO	15,312	5,500,000	429,000
H_2	31,260	11,220,000	62,520
N_2	6,690	2,400,000	187,200
CH_4	2,360	847,000	37,800
H_2O	144	51,700	2,600
Total	73,518	26,388,700	1,500,120

No information is available on the effect of CH_4, H_2O, etc. in the recycle on the conversion. Therefore this CH_4 and H_2O will be considered to pass through the reactor unchanged for purpose of material and energy balances.

Feed To Reactor Composition:

Reactor Feed Constituents	Mols/hr Feed & Recycle & Regen Cal	Total	Standard ft^3/hr	#'s/hr
CO_2	197.3 + 17,750	= 17,947.3	6,445,000	289,680
N_2	290.2 + 6,690	= 6,980.2	2,505,000	195,330
H_2	14,500.0 + 31,260	= 45,760.0	17,100,000	91,520
CO	7,630.0 + 15,312	= 22,942.0	8,240,000	642,500
CH_4	22.0 + 2,360	= 2,382.0	855,000	38,152
H_2O	1,190 + 144 + 500	= 1,835.0	679,000	33,000
		97,348.5	35,110,000	1,789,212

Reactor Product Composition:

Reactor Product Constituents	Mols/hr Feed + Product	Total	St'd ft^3/hr	#'s/hr
CO_2	17,947.3 + 575 =	18,522.3	6,650,000	817,000
N_2	6,980.2 ---- =	6,980.2	2,505,000	195,330
H_2	45,760.0 - 13,127 =	32,633.0	11,710,000	65,266
CO	22,942.0 - 7,437 =	15,505.0	5,560,000	445,000
CH_4	2,382.0 - 22 =	2,360.0	847,500	37,752
H_2O	1,835.0 + 5,783 =	7,618.0	2,740,000	128,200
C	76.3=	76.3		915
Product	+ 6,360 =	6,360.0	2,282,000	91,664
		90,054.8	29,454,500	1,789,212

Product	Wt %	Mol/100# Product	Avg Mol Wt	Mols 1 hr
$C_1 - C_2$	16	.889	18	3,000
$C_3 - C_4$				1,298
C_nH_{2n+2}	5.7	.112	51	
C_nH_{2n}	13.3	.272	49	
Gasoline(C_5-C_{10})				1,340
C_nH_{2n+2}	10.5	.098	107	
C_nH_{2n}	31.5	.300	105	
Diesel Oil (C_{10}-C_{18})				197
C_nH_{2n+2}	4.8	.023	205	
C_nH_{2n}	7.2	.036	203	
Wax (C_{18} +)	1.0	.00354	282	11.8
Water-Soluble Chem.				
Alcohols $C_nH_{2n-2}O$	3.76	.0723	52	245
Aldehydes $C_nH_{2n}O$	.75	.0144	52	49
Ketones $C_nH_{2n}O$	.78	.0130	60	44
Acids	1.95	.0305	64	103
Oil Soluble Chem.	2.76	.0328	116	80
		1.887		

REACTOR SELECTION

B. DESIGN

The design of the reactor has been very much of an approximation at the best. This has been caused by the scarcity of information about fluidized beds using iron catalyst for the Fischer-Tropsch process. No correlation between the design and operating variables of the reactor and the products obtained was available. The information in the literature led to only a few generalizations. It is these generalizations which have been used to select a process and design the reactor. The results are at best only a rough estimate of the results that should be used.

When a product had been established by a series of reasoned estimates from the literature as described previously, a material balance was made on the reactor. This led to an estimate of the fresh feed necessary and the composition of the fresh feed. The value of 3 to 1 for the ratio

of fresh feed to recycle having been mentioned in Bureau of Mines Technical Paper 209, in reports on the Carthage Hydrocal Plant and the Standard Plant this was chosen as the one to be used. From a survey of the temperatures and pressures mentioned in the literature, the operating conditions of the reactor were set at 600°F and 300 psia. In consultation with the products group, the recycle stream temperature was set at 85°F. The synthesis gas group set the fresh feed stream at 200°F. A heat balance around the reactor now revealed that 153×10^6 Btu/hr would have to be removed from the reactor.

Before the actual design of the reactor could begin, it was necessary to have some idea of the proper space velocity to use. References to actual space velocities in the literature were scarce and these ranged from around 50 ft.^{3}/ft.^{3}/hr. to 1000 ft.^{3}/ft.^{3}/hr. A value of 650 ft.^{3}/ft.^{3}/hr was finally selected as being in fairly good agreement

REACTOR SELECTION

with the space velocity Carthage Hydrocal was apparently using. The L/D ratio of the reactors was fixed at 3.3 to 1 as being a good one for fluidization efficiency according to an article by L.T. WORK in I.E.C. page 1099, JUNE, 1949 VOL. 41

Calculations:

Space Velocity = 650 ft.^{3} fresh gas / ft.^{3} catalyst / hr.

Fresh Feed = 8,550,000 ft^3 / hr.

Reactor Volume = 8,550,000 ÷ 650 = 13,150 ft.^{3}

According to the literature ( ) and picture in Fortune Carthage Hydrocal is using 4 reactors

$\frac{13,150}{4}$ = 3295 ft^3 catalyst space / reactor

Volume = $\frac{\pi D^2}{4} L = 2.59 D^3$

Diameter of Reactor = $\left[\frac{3295}{2.59}\right]^{1/3}$ = 10.9 ft.

Height of Reactor, L = 3.3 D = 35.8 ft.

It has been seen that 153×10^6 Btu / hr. had to be removed from the reactor to maintain the temperature at 600° F. This heat is to be removed

B. DESIGN (CONT.)

by generating steam in tubes located in the catalyst bed. The requirements of the plant for high pressure steam agreeing with references in the literature to the generation of 680 psia steam led to the decision to generate steam at this pressure.

A steam side coefficient of 100 Btu/hr.-ft^2-°F. was chosen from Chem. Eng. Hdbk. From Robert Toomey's Ph.D. Thesis ( ), the value of the catalyst side coefficient of 80 Btu/hr.-ft^2-°F was selected as being a reasonable one to expect.

Overall Coef.

$$\frac{1}{U_o} = \frac{1}{h_{steam}} + \frac{1}{h_{cat.}} = \frac{1}{100} + \frac{1}{80}$$

$$U_o = 60 \text{ Btu/hr-ft}^2\text{-°F.}$$

The steam pressure of 680 psia equivalent to 500°F fixed the temperature difference at 100°F.

$$\text{Surface Area} = \frac{q}{U\Delta t} = \frac{153 \times 10^6}{60 \times 100} = 25.2 \times 10^3 \text{ ft}^2$$

$$\text{Area/reactor} = \frac{25200}{4} = 6300 \text{ ft.}^2$$

REACTOR SELECTION

In class consultation the size of the steam tubes was set at 2" inch O.D. 7 BWG

~~2 inch~~ 2 inch - 7 BWG. = .4294 ft^2 surface / ft of length

$$\frac{6.300}{.4294} = 14,680 \text{ ft. tubing / reactor}$$

Using 35 ft. hairpin tubes

Hairpin tubes / reactor = $\frac{14,680}{70} = 210$

Multiclones for separating entrained catalyst were designed according to Section — of 3rd Edition of Chem. Engr. Hobbs.

95% of 5 μ diam. particles to be removed

9 in. diam. multiclones

$$\frac{D_x}{D_{min}} = \sqrt{x(2-x)^{.5}} = \sqrt{.95(1.05)^{1/2}} = .989$$

$$D_{min} = \frac{5}{.989} = 5.05\,\mu$$

$$D_{min} = \left[\frac{9\mu b}{\pi V N(\rho_p - \rho_g)}\right]^{1/2}$$

$b = D/4 = 2.25$ in $= .1875$ ft

μ = viscosity $= .0714 \frac{\#}{hr\text{-}ft}$

$N = 5$ $\quad \rho_g = .735 \#/ft^3$

$\rho_p = 490 \#/ft^3$

$$\frac{5.05}{30.5\times10^4} = \left[\frac{9 \times 7.14\times10^{-2} \times .1875}{\pi \times V \times 5 \times (490-.7)}\right]^{1/2}$$

REACTOR SELECTION

DESIGN (CONT.)

Velocity, V = 58,300 ft/hr = 16.2 ft/sec.

Volume of Gas = $31{,}380{,}000 \frac{ft^3}{hr} \times \frac{14.7}{300} \times \frac{1060}{492} \times \frac{1}{60} = 55{,}000\ ft^3/min.$

Volume / reactor = 13,750 ft^3/min.

Cross Sectional Area of Inlet / cyclone

$$A = \frac{\frac{D}{2} \times \frac{D}{4}}{144} = \frac{4.5 \times 2.25}{144} = .0703\ ft^2$$

Number of Cyclones: $\frac{13{,}750}{60} \times \frac{1}{.0703} \times \frac{1}{16.2} = 201$ cyclones

Height of cyclone = 4D = 4 × 3/4 = 3 FT.

Height of Reactor:

Catalyst Space + Header + Cyclone + Clearance =
36 ft. + 2 ft. + 3 ft + 1 = 42 ft.

Dimensions of Reactor = 42 FT. × 11 FT.

Wall Thickness of Reactor

Designed on basis of API-ASME Code

$$t = \frac{pD}{2Se - p} + C$$

t = wall thickness
D = inside diam. of vessel
p = max. allowable working pressure
S = max. allowable tensile strength
e = efficiency of longitudinal joints
C = corrosion allowance

B. DESIGN (CONT.)

S_u = Ultimate Tensile Strength of Steel = 60,000 psi

$S = 60{,}000 \times .25 = 15{,}000$ psi

$e = .80$ for this type vessel

$p = 415$ psig - safety factor

$c = .25$ in. for abrasion and possible case hardening

$$t = \frac{415 \times 11 \times 12}{24{,}000 - 415} + .25 = 2.57 \text{ inches}$$

REACTOR SELECTION

C. COST ESTIMATE

1) Cost of Shell: based on weight of steel in shell and cost in Bliss with factor of 1.75 to bring cost to date

Area of Heads = $2 \times 1.09 D^2$ = 264 ft^2
Area of Sides = $\pi \times 11 \times 42$ = 1450 ft^2

Total Area = 1714 ft^2

Weight of Steel = $1714 \times \frac{2.57}{12} \times 7.9 \times 62.4$ = 181,000 lb/reactor

Cost = 181,000 × .10 × 1.75 = $31,700 / reactor

Cost of Heat Exchange Surface

Based on method of Bliss. Chrome-manganese steel used to resist abrasion and temperature.

Cost of Tubes: $3.30/ft.

14,680 × $3.30 = $48,500 /reactor

Cost of Header: Based on surface area

45 ¢/ft^2, Factor of 4.0 for Chrome-manganese

6,300 × .45 × 4.0 × 1.75 = $19,800 /reactor

C. COST ESTIMATE (CONT.)

Cost of Cyclone

Estimated according to Ch. Eng. Hobbs 3rd Edit.

50 ¢/ft³ gas/min. 13,750 ft³ gas/min.

13,750 x .50 = $6,900

Total Cost of Reactor = $31,700 + $48,500 + $19,500 + $6,900 =
$106,750

Regenerator Design

Due to a definite lack of information in the literature as to the regeneration of Fischer-Tropsh iron catalysts, it was necessary to base the regenerator design on procedures followed in the regeneration of petroleum cracking catalysts. The qualification for this is the fact that the object in both cases is to remove carbon depositions from the catalyst. The F-T regeneration is somewhat more complicated since wax must also be removed. However both may be removed by the same process, namely, by burning with air.

The allowable fouling of the catalyst was set at 1 percent by weight while the deposition of carbon and wax was estimated to be 1835 pounds per hour. Calculations based on this and the total amount of catalyst present in the reactors indicated that 183,300 pounds per hour of catalyst must be completely regenerated. The amount of air necessary for the burning was determined allowing a 20 percent excess to insure complete regeneration. A flame temperature calculation gave a regenerator temperature of approximately 1200°F.

To determine the diameter of the regenerator, a Stokes Low Settling velocity of 3.14 ft. per sec. for the catalyst was calculated for the 33 micron particle and this was raised to 5 ft per sec to allow for hindered settling. Based on this velocity the diameter of the regenerator is 10 ft. As cracking catalyst regenerators appear to have the same height as diameter, the height of the regenerator was taken to be 10 ft. The shell is $\frac{1}{4}$ in. steel with a fire brick lining to withstand the high temperature.

It was not considered feasible to generate steam utilizing the heat in the regenerator gases as the low heat transfer coefficients would require too large a heat exchanger to perform the operation.

It was desired to recover 95 per cent of particles 6 micron in diameter. To accomplish this, 250 multiclones are placed in the top of the shell in an inclined, radial position.

The spent catalyst is carried up the riser to the regenerator by the regenerating air. As this is at low pressure, slightly over atmospheric, a screw-type valve is used at the entrance to the riser from each reactor to control the flow of catalyst at 300 psi in the reactor to the riser at lower pressure. The air is taken from the surroundings and blown by a centrifugal blower up the riser.

To return the regenerated catalyst to the reactor, a stand pipe is used which projects into the catalyst bed nearly to the surface. At the lower end of the standpipe or downcomer, a screw-type valve ejects the catalyst into the fresh feed line. Hydrogen is allowed to enter the downcomer to keep the catalyst from packing or bridging and also serves to reduce any of the catalyst which was oxidized during regeneration. Also, new catalyst to replace the 1 percent loss per month enters this downcomer.

Regenerator Material Balance

In

Spent catalyst:

Fe	183,300 #/hr.	
C	916	
Wax	919	
Total		185,135 #/hr.
Air for regeneration		30,220
Total In		215,355 #/hr.

Out

Regenerated catalyst		183,300 #/hr.
Regenerator gases:		
CO_2	6,230 #/hr.	
H_2O	1,232 #/hr.	
O_2	1,413	
N_2	23,180	
Total		32,055
Total Out		215,355 #/hr.

Regenerator Heat Balance

In

Enthalpy of catalyst & air stream to regenerator:*

Constituent	mols/hr.	c_p	Δt	Btu/hr.
Fe	3280	8.49	(600-32)	15,800,000
O_2	220	6.90	(60-32)	42,500
N_2	826	6.90	(600-32)	159,700

Total — — — — — — — — 16,020,200 Btu/hr.

Heat of reaction minus heat losses — — — — 25,865,800

Total heat in — — — — — — — — 41,886,000 Btu/hr.

Out

Enthalpy of regenerator gases from regenerator:

Constituent	mols/hr.	c_p	Δt	Btu/hr.
CO_2	141.5	10.80	(1200-32)	1,782,000
H_2O	68.5	8.60	"	688,000
O_2	44.2	6.90	"	356,000
N_2	826.0	6.90	"	6,660,000

Total — — — — — — — — 9,486,000 Btu/hr.

Enthalpy of catalyst from regenerator

Constituent	mols/hr.	c_p	Δt	Btu/hr.
Fe	3280	8.49	(1200-32)	32,400,000

Total — — — — — — — — — 32,400,000

Total heat out — — — — — — — — 41,886,000 Btu/hr.

* Heat content of wax and carbon on iron neglected as small

Regenerator Design Calculations

I. Catalyst Circulation

Mols carbon deposited per 100 mols of feed — — — — = 0.32

Mols per hour of feed gas — — — — — — — — = 23,850.5

Mols carbon deposited per hour = $\frac{0.32(23,850.5)}{100}$ — — — — — — = 76.30

Average wax constituent taken as $C_{30}H_{42}$

#$C_{30}H_{42}$ deposited per 100 mols of feed — — — — — = 3.85

Molecular weight — — — — — — — — = 282

Mols $C_{30}H_{42}$ deposited per 100 mols of feed = $\frac{3.85}{282}$ — — — — = 0.01362

Mols $C_{30}H_{42}$ deposited per hour = $\frac{0.01367(23,850.5)}{100}$ — — — = 3.26

#Carbon deposited per hour = 76.30 (12.01) — — — — — = 916.0

$C_{30}H_{42}$ deposited per hour = 3.26 (282) — — — — — = 919.0

Iron catalyst data [I & E.C., 41, 1100 (1949)]

Diameter: 33 μ

Particle density: 490 #/ft³

Bulk density: 134 #/ft³

Maximum aerated bed density: 100.0 #/ft³

Reactor bed volume, ft³ — — — — — — — — — = 3300

Total reactor bed volume ft³ = 4 (3300) — — — — — = 13,200

Wt. of catalyst, # = 13,200 (100.0) — — — — — — = 1,320,000

Deposition, # per # catalyst = $\frac{916+919}{1,320,000}$ — — — — — = 0.00139

Allowable fouling of catalyst, % — — — — — — = 1

Catalyst circulated for regeneration, #/hr. per # catalyst = $\frac{0.00139}{0.01}$ = 0.139

Total catalyst recirculated for regeneration, #/h. = 0.139 (1,320,000) = 183,300

Mols catalyst recirculated per hour = $\frac{183,300}{55.84}$ — — — — = 3280

II. Temperature in Regenerator

Thermodynamic Data:

Heats of combustion at 25°C, 1 atm., to $CO_{2(g)}$ and $H_2O_{(g)}$

$C_{30}H_{42}$: 18,857 Btu/# (Perry)

C : 94,000 Btu/# (Daniels)

Heat capacities

Fe : 8.49 Btu/#mol °F
C : 4.65 Btu/#mol °F
$C_{30}H_{42}$: 141 Btu/#mol °F
} Perry

Heat capacities (cont)

$$C_p = a + bT - cT^2 \quad Btu/\#_{mol}\,°F$$

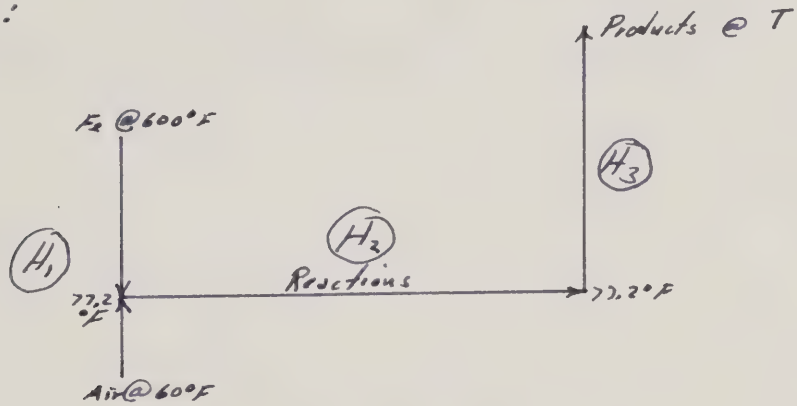
	a	b	c	
$CO_{2(g)}$	10.34	0.00152	—	Clarke, p. 103
$H_2O_{(g)}$	6.95	0.0180	0.100×10^{-6}	
$N_{2(g)}$	6.45	0.00080	0.025×10^{-6}	
$O_{2(g)}$	6.10	0.0018	0.314×10^{-6}	

Reactions:

$$C_{20}H_{42(\ell)} + 30\tfrac{1}{2}\,O_{2(g)} \longrightarrow 20\,CO_{2(g)} + 21\,H_2O_{(g)}$$

$$C + O_{2(g)} \longrightarrow CO_{2(g)}$$

Heat Balance:



$$H_1 + H_2 = H_3$$

Material Balance:

Mols $C_{20}H_{42}$ per hour — — — — — — — = 3.26

Theoretical use of O_2 for combustion of $C_{20}H_{42}$, mols/hr = 3.26 (30.5) — — — = 99.5

Mols Carbon per hour — — — — — — — = 76.3

Theoretical use of O_2 for combustion of Carbon, mols/hr = 76.3 (1) — — = 76.3

Total Theoretical O_2, mols/hr = 99.5 + 76.3 — — — — = 175.8

Composition of air: 21% O_2, 79% N_2

Theoretical air needed, mols/hr = $\frac{175.8}{0.21}$ — — — — — — — = 837.0

Will use 20% excess air for combustion

Actual air needed, mols per hour = $\frac{837.0}{0.80}$ = — — — — = 1046.0

Actual O_2 needed, mols per hour = 1046.0 (0.21) = — — — = 220.0

Actual N_2 needed, mols per hour = 1046.0 (0.79) — — — — = 826.0

CO_2 produced, mols per hour = 3.26 (20) + 76.3 (1.0) — — — — = 141.5

H_2O produced, mols per hour = 3.26 (21) — — — — — = 68.5

Reactants		Products	
$C_{20}H_{42}$:	3.26	CO_2:	141.5
C:	76.30	H_2O:	68.5
O_2:	220.00	O_2:	44.2
N_2:	826.00	N_2:	826.0
Fe:	3,280.00	Fe:	3,280.0

H_1 Btu/hr.

$C_{20}H_{42}$: $3.20\,(141)(600-77.2)$ — — — — — — = 240,300

C : $76.3\,(4.65)(600-77.2)$ — — — — — — = 185,600

Fe : $3280\,(8.49)(600-77.2)$ — — — — — — = 14,580,000

O_2 : $220.0\int_{537.2}^{500}(6.10+0.0018T-0.314\times10^{-6}T^2)\,dT$

N_2 : $8260\int_{537.2}^{500}(6.46+0.00880T-0.025\times10^{-6}T^2)\,dT$

O_2+N_2 := — — — — — — — = − 124,000

H_1 Btu/Hr — — — — — — — = 14,881,900

H_2, Btu/hr

Heat of reaction of $C_{20}H_{42}$ = $3.26(282)(18,857)$ = — — — — = 17,330,000

Heat of reaction of Carbon = $76.3\,(169,200)$ — — — — — — = 12,900,000

H_2 Btu/hr — — — — — — — — — = 30,230,000

H_3 Btu/hr

$$H_3 = 141.5\underbrace{\int_{537.2}^{T}(10.34+0.00152T)\,dT}_{CO_2} + 68.5\underbrace{\int_{537.2}^{T}(6.95+0.0180T-0.100\times10^{-6})\,dT}_{H_2O}$$

$$+ 44.2\underbrace{\int_{537.2}^{T}(6.10+0.0018T-0.314\times10^{-6}T^2)\,dT}_{O_2} + 826.0\underbrace{\int_{537.2}^{T}(6.45+0.0088T-0.025\times10^{-6}T^2)\,dT}_{N_2}$$

$$+ 3280.0\,(8.49)(T-537.2)$$

Solving $H_1+H_2=H_3$, integrating and solving for T,

T in regenerator = — — — — — — — 1272°F

Due to heat losses, the temperature of the regenerator will be taken as 1200°F and subsequent calculations will be based on this value.

III Size of regenerator

Fluidizing velocity will be based on Stokes law settling and the value obtained scaled up to allow for interference of adjacent particles.

Air for regeneration (14.7 psia, 60°F), mols/hr - - - - - - = 1046

Volume of air to riser, ft^3 per hour $= 1046(359)\frac{520}{492}$ - - - - - = 397,000

Gases from regenerator (14.7 psia, 1200°F), mols per hr. - - - - - = 1080.2

Volum. of gases from regenerator (14.7 psia, 1200°F) $ft^3/hr. = 1080.2(359)\frac{1660}{492} = 1,310,000$

$ft^3/sec = \frac{1,310,000}{3600}$ - - = 364.0

Volum of air to regenerator (14.7 psia, 1200°F) $ft^3/h = 1046(359)\frac{1660}{492} = 1,267,000$

$ft^3/sec = \frac{1,267,000}{3600}$ - - - = 352.0

Reactants (gases)

O_2 : 7,040 #/hr.

N_2 : 23,180

30,220 #/hr.

Products (gases

CO_2 : 6,230 #/h.

H_2O : 1,232

O_2 : 1,413

N_2 : 23,180

32,055 #/hr.

$\rho_{air} = \frac{30,220}{1,267,000} = 0.0239$ #/ft3

$\rho_{gases} = \frac{32,055}{1,310,000} = 0.0245$ #/ft3

ρ_{ave}, #/ft3 $= \frac{0.0239+0.0245}{2}$ - - - - - - - = 0.0242

μ_{air}, centipoises - - - - - - - - = 0.04

Dia. of particle $= 33\times10^{-4}$ cm $= \frac{33\times10^{-4}}{2.54\times12}$ ft - - - - - - = 1.082×10^{-4}

g, ft/sec² - - - - - - - - = 32.2

Density of particle, #/ft3 - - - - - - - - = 490

$$C_D R_e^2 = \frac{4 g D^3 \rho_g (\rho_p - \rho_g)}{3\mu^2}$$

$$C_D R_e^2 = \frac{4(32.2)(1.082\times10^{-4})^3(0.0242)(490)}{3(0.04\times0.000672)^2}$$ - - - - - - = 0.875

$$Re = \frac{Du\rho}{\mu} = \frac{(1.082\times10^{-4})\,u\,(0.02373)}{0.04\times0.000672}\ (\text{Perry}) =$$ - - - - - - = 0.03

u, ft/sec - - - - - - - - - - - - = 3.14

Take fluidizing velocity to be <u>5 ft/sec</u> to allow for interference of other particles.

Cross section of regenerator, $ft^2 = \frac{364.0}{5}$ = — — — — = 72.8

Radius of regenerator, $ft = \sqrt{\frac{72.8}{3.1416}}$ = — — — — = 4.71

Take diameter of shell, ft — — — — 10

Take height of regenerator, patterned after petroleum catalyst regenerator, ft = 10

Area of regenerator shell, $ft^2 = 3.1416(10)^2 + 218$ = — — — = 532

No. of fire brick for lining (0.1875 ft^2 per brick) $= \frac{532}{0.1875}$ = — — — = 2,840

Thickness of steel shell, in = — — — — — 2.5

Volume of steel in shell, $ft^3 = \frac{532(2.5)}{12}$ — — — — — = 11.1

IV. Cyclones

Multicones to be used. To be set radially and inclined in the top of the regenerator. Designed according to Sect. 15 of Chem. Eng's Hdbk. 9" multicones are to be used. 95 percent of 5 μ particles to be removed.

$\frac{D_x}{D_{min}} = \sqrt{x(2-x)^{0.5}} = \sqrt{0.95(2.05)^{0.5}}$ — — — — — = 0.989

D_{min}, microns, $= \frac{5}{0.989}$ — — — — — — = 5.05

$D_{min} = \sqrt{\frac{9\mu b}{\pi V N(\rho_p - \rho_g)}}$

$b = D/4$ — — — — — = 0.1875

μ, lb/hr ft — — — — — — — = 0.0956

N — — — — — — — 5

ρ_p, lb/ft³ — — — — — — — = 490

ρ_g, lb/ft³ — — — — — — — = 0.0242

$$\frac{5.05}{30 \times 10^{-4}} = \left[\frac{9(0.0956)(0.1875)}{\pi(V)(5)(490)}\right]^{1/2}$$

V, ft/sec $= \frac{(30.5 \times 10^{-4})^2(9)(0.0956)(0.1875)}{(5.05)^2(\pi)(5)(490)}$ — — — — = 21.2

Gas volume, scf/min $= \frac{1080.2\ (359)}{60}$ — — — — — = 6,470

Gas volume (1200°F, 142 psia) ft³/min $= 6470 \frac{1660}{492}$ — — — — = 21,800

Area of inlet per multicone $= \frac{\frac{D}{2} \times \frac{D}{4}}{144} = \frac{4.5 \times 2.25}{144}$ ft² — — — = 0.0703

No. of multicones necessary $= \frac{21,800}{60} \times \frac{1}{0.0703} \times \frac{1}{21.2}$ = — — — — = 244

No. of multicones actually used — — — — = 250

Height of multicones $= 4D = 4(3/4)$ — — — — = 3 ft.

V Regenerator costs

1. Shell

Weight of steel, # — = 5,460

Cost of steel, fabricated/# — = $0.30

Cost of shell = 5,470(0.30) — = $1640

2. Shell lining

No. of bricks — 2,840

Cost/1000 bricks — $75

Cost of lining = 2.84(75) — = $213

3. Multicloncs

No. of multicloncs — 350

Cost/ft³ of gas handled/min (Perry) — = $0.50

Volume of gas handled, ft³/min — 21,800

Cost of multicloncs = 21,800 (0.50) — = $1100

Total Regenerator Cost — = $2,953

Regerator and Reactor Piping Design

I Design

1. Riser

It was felt that a reasonable diameter for the riser would be 30 inches. The velocity in the riser is based on this figure

Velocity, ft/sec = $\frac{397{,}000\,(12)^2}{(3600)(3.14)(15)^2}$ = 22.5

Thickness, in = 3/8

Length, ft = 65

The riser will be constructed of chrome-manganese steel to withstand abrasive action of catalyst

2. Downcomer.

Downcomer will be 1 ft in diameter. This was selected as sufficient to allow for partial fluidization of the regenerated catalyst and to minimize bridging. It will be constructed of chrome manganese steel.

Thickness, in = 3/8

Length, ft = 64

The other pipes were designed so as to have the gases at a velocity of the same order of magnitude as in the riser. Lengths are based on extention of 50 ft. past equipment except where connecting pieces of equipment in the reactor section.

3. Fresh feed pipe

Diameter, ft = 2.5

Flow, ft³/sec = 144.2

Velocity, ft/sec = $\frac{144.2}{3.1416(1.25)^2}$ = 29.4

Length, ft. = 50

Thickness, in = 1/2

4. Recycle Pipe

Diameter, ft = 4

Flow, ft³/sec = 357

Velocity = $\frac{357}{3.1416(2)^2}$ = 29.4

Length, ft = 70

Thickness, in = 7/8

5. Total feed pipe

Diameter, ft — — — — — — — = 5

Flow, ft^3/sec — — — — — — = 501

Velocity, ft/sec $= \frac{501}{3.1416(2.5)^2}$ — — — — — = 25.5

Length, ft — — — — — — = 15

Thickness, in — — — — — — — = 1

6. Product pipe

Diameter, ft — — — — — — — — .5

Flow, ft^3/sec — — — — — = 917

Velocity, ft/sec $= \frac{917}{3.1416(2.5)^2}$ — — — — — — = 46.7

Thickness, in — — — — — — — 1

II Costs of Piping

1. Riser

Cost/# — — — — — — $2.00

Cost $= 3.1416(2.5)(3/8)(\frac{1}{12})(65)(7.9)(62.4)(2.00)$ — — — — — = $15,740

2. Downcomer

Cost/# — — — — — — $2.00

Cost $= 3.1416(1)(3/8)(\frac{1}{12})(64)(7.9)(62.4)(2.00)$ — — — — — = $6,200

3. Fresh feed pipe

Cost/# — — — — — $0.25

Cost $= 3.1416(2.5)(\frac{1}{2})(\frac{1}{12})(50)(7.8)(62.4)(0.25)$ — — — — = $1,990

4. Recycle pipe

Cost/# — — — — — — $0.25

Cost $= 3.1416(4)(3/8)(\frac{1}{12})(70)(7.8)(62.4)(0.25)$ — — — — — = $7,800

5. Total feed pipe

Cost/# — — — — — — $0.25

Cost $= 3.1416(5)(\frac{1}{12})(15)(7.8)(62.4)(0.25)$ — — — — — = $2,490

6. Product pipe

Cost/# — — — — — — — $0.25

Cost $= 3.1416(5)(\frac{1}{12})(50)(7.8)(62.4)(0.25)$ — — — — — = $7,960

Total piping cost — — — — — — $42,180

COMPRESSORS

A. Fresh Feed

Fresh Feed is received from the Synthesis Gas plant at 275 psia. To flow through the reactors it must be compressed to 325 psia. Calculations are based on theoretical adiabatic compression using factors of 65% for the compressor efficiency and 75% for the turbine efficiency.

$$\text{Theoretical H.P.} = \frac{k\, p_1 V_1}{33,000(k-1)}\left[\left(\frac{p_2}{p_1}\right)^{\frac{k-1}{k}} - 1\right]$$

$$V_1 = 8,541,700 \times \frac{14.7}{275} \times \frac{660}{492} = 612,000 \text{ ft}^3/\text{hr.}$$

$k = 1.3$

$$\text{Theoretical H.P.} = \frac{1.30 \times 275 \times 612,000}{33,000 \times 0.30 \times 60}\left[\left(\frac{325}{275}\right)^{.23} - 1\right] = 14.4 \text{ H.P.}$$

$$\text{Horsepower required} = \frac{14.4}{.65 \times .75} = 29.6 \text{ H.P.} \doteq 78,400 \text{ Btu/hr.}$$

Cost: Villbrandt ; \$100/HP. $100 \times 30 =$ \$3000

B. Regeneration air

The air used to burn off the deposits on the catalyst and fluidize the regeneration is required at 30 psig. 397,000 ft³/hr at STP required

B. Regeneration Air (cont.)

$$\text{Theoretical HP.} = \frac{k\, p_1 V_1}{33{,}000(k-1)}\left[\left(\frac{p_2}{p_1}\right)^{\frac{k-1}{k}} - 1\right]$$

$$\text{Theoretical HP.} = \frac{1.4 \times 15 \times 397{,}000}{33{,}000 \times 60 \times 0.40}\left[\left(\frac{45}{15}\right)^{.286} - 1\right] = 4.0$$

$$\text{Actual HP.} = \frac{4.0}{.65 \times .75} = 8.2 \approx 20{,}900 \text{ Btu/hr.}$$

Cost: Nilbrandt: \$100/HP. $100 \times 8.2 =$ \$820

INSTRUMENTATION

The accurate control of the reactor temperature and pressure is absolutely essential to the successful operation of the plant. The increase in methane formation with increasing temperature and the increase in wax formation with increasing pressure are two factors which illustrate the necessity of good control. The regenerator temperature and pressure while not so critical still have a material effect on the efficiency of regeneration and catalyst life.

The temperature of the reactor can be controlled in the first place by regulating the amount of fresh feed. For this purpose at point 1 [point numbers refer to points indicated on the flow sheet diagram] a flow recorder and controller will be installed. At point 1 will be also a pressure recorder and controller to control compressor C_1 to maintain the pressure constant. At point 2 will be a flow recorder on the recycle stream. The temperature and pressure of the total flow to each reactor will be recorded and controlled at points 5_1, 5_2, 5_3, 5_4.

The second important way to control the temperature of the reactor is by varying the amount

of steam and the condition of the steam formed in the reactor. Thus lowering the steam side pressure will have the effect of increasing the temperature difference and thus increasing the rate of removal of heat from the reactor. For this purpose at points 6_1, 6_2, 6_3, and 6_4 will be flow recorder and controllers on the feed water to each reactor. At points 11_1, 11_2, 11_3, and 11_4 will be pressure recorders and controllers to maintain the desired steam pressure.

From the material balance, it can be seen that a volume contraction occurs during the reaction. Any increase in methane, carbon dioxide, or water formation will decrease this contraction. Thus, maintaining the correct temperature will help prevent pressure increases. The reactor pressure will be recorded in each reactor at points 12_1, 12_2, 12_3, and 12_4. The operating pressure can be adjusted by varying the output pressure of the fresh feed and recycle steam compressors.

The flow of catalyst to the regenerators will be controlled by the amount of deposit collecting on the catalyst. This will be determined by control

samples taken at points 13_1, 13_2, 13_3, and 13_4. The flow of catalyst to the regenerator will be controlled by a screw feed. This flow will be recorded at points 13_1, 13_2, 13_3, and 13_4 and will control the amount of air admitted at point 4 for fluidization. The temperature of the regenerator as indicated by the temperature recorder at point 10 will control the excess air feed at point 4. If necessary the amount of catalyst could also be cut down if the regenerator temperature was too high. The condition of the regenerated catalyst will be determined at sampling point 3 just before it is fed by a screw feed into the reactor feed stream. The flow of regenerated catalyst will also be recorded at point 3. The amount of fresh catalyst will be recorded and controlled at point 14.

The temperature of the exit gases from the regenerator will be recorded at point 3. The amount and temperature of the product stream from each reactor will be recorded at points 7_1, 7_2, 7_3 and 7_4. The temperature throughout the reactor beds will be recorded at points 9_1, 9_2, 9_3, and 9_4.

INSTRUMENTATION

COST ESTIMATE

	Number
Total Instruments Required	23
Flow Recorder and Controller	16
Pressure Recorders and Controller	5
Pressure Recorder	15
Temperature Recorder, Thermocouple	5
Multipoint Temperature Recorder	

Instrument Costs from Villbrandt - 3rd Edition

Flow Recorders and Controllers 23 @ $500	$11,500
Pressure Recorders 16 @ $150	2400
Pressure Recorder and Controller 5 @ $350	1750
Temperature Recorder 15 @ $150	2250
Multipoint Temperature Recorder 5 @ $800	4000
Total	$21,950

Cost Estimate

I. Reactor:

A. Shell	$31,700
B. Tubes	$48,500
C. Headers	$19,200
D. Multiclones	$6,900
	$106,300

Four Reactors required: --- Total -- $425,200

II. Regenerator:

A. Shell	$1,640
B. Fire brick Lining	$213
C. Multiclones	$1,100
	$2,953

One Regenerator required: ---- $2,953

III. Pipes:

A. Fresh Feed Pipe	$1,990
B. Total Feed Pipe	$2,490
C. Riser to Regenerator	$15,740
D. Standpipe	$6,200
E. Recycle Pipe	$7,800
F. Product Pipe	$7,960
Total	$42,180

Cost Estimate -(cont.)

IV. Catalyst: -- 100 tons @ $105/ton --- $10,000

V. Compressors:

A. Compressor or Booster on Fresh Feed -- $3,000

B. Air Blower for Regenerator ---- $820

Total: $3,820

VI. Instrumentation ---------- $21,900

Total Equipment Cost (I, II, III, IV, V, VI) --- $512,070

The large weight of the equipment, long pipes involved and the complexity of the reactors will make erection cost high. Therefore erection costs will be estimated at 100 % of equipment costs:

VII - Installation Cost:

100 % of Equipment Cost ---- $512,070

Total Installed Equipment Cost ---- $1,024,140

VIII. - Depreciation - 10 % of Equipment Cost = $51,200/yr.

IX. - Taxes, Rent, Insurance - 10% of Equipment Cost --- $51,200/yr.

X - Labor - 2 Men @ $3.00/hr. ---------- $52,500/yr.

Cost Estimate -(cont.)

XI - Fresh Catalyst - - - - - - - - - - $1,200/yr.

Operating Costs (exclusive of materials & power

VIII & IX & X & XI - - - - $156,114/yr.

FLOW SHEET

On the large flow sheet the letters refer to streams in the plant and the numbers to control points in the plant. Subscripts are used to indicate which reactor is being referred to. The temperature, pressure, mass flow rate, and enthalpy of each stream will be given below. For more detailed information on each stream reference should be made to the appropriate section of the report.

Fresh Feed - F_B

Temperature - 200°F

Pressure - 325 psia.

Mass Flow Rate - 281,092 #/hr.

Enthalpy - 45.8×10^6 Btu/hr.

Fresh Feed + Regenerated Catalyst + Recycle - F_D

Temperature - 171°F

Pressure - 325 psia.

Mass Flow Rate - 1,789,212 #/hr.

Enthalpy - 118.43×10^6 Btu./hr.

FLOW SHEET

Feed to Reactor - $f_1 = f_2 = f_3 = f_4$

Temperature - 171°F

Pressure - 325 psia

Mass Flow Rate - 447,303 lbs./hr.

Enthalpy - 29.61×10^6 Btu/hr.

Product Stream - $p_1 = p_2 = p_3 = p_4$

Temperature - 600°F

Pressure - 300 psia

Mass Flow Rate - 447,303 lbs./hr.

Enthalpy - 157.03×10^6 Btu/hr.

Combined Product Stream - P

Temperature - 600°F

Pressure - 300 psia

Mass Flow Rate - 1,789,212 lbs./hr.

Enthalpy - 628.10×10^6 Btu/hr.

Recycle Stream - R

Temperature - 85°F

Pressure - 325 psia

Mass Flow Rate - 1,500,120 lbs/hr.

Enthalpy - 32.65×10^6 Btu/hr.

FLOW SHEET

Catalyst to Regenerator - $a_1 = a_2 = a_3 = a_4$

Temperature	-	600° F
Pressure	-	325 psia
Mass Flow Rate	-	46,284 lb./hr.
Enthalpy	-	3.95×10^6 Btu./hr.

Catalyst to Regenerator, Combined Stream - A_c

Temperature	-	1000° F
Pressure	-	30 psia
Mass Flow Rate	-	215,355 lb./hr.
Enthalpy	-	41.89×10^6 Btu./hr.

Regenerator Stack Gases - D

Temperature	-	1200° F
Pressure	-	15 psia
Mass Flow Rate	-	32,055 lb/hr.
Enthalpy	-	9.486×10^6 Btu/hr.

Regenerated Catalyst + Fresh Makeup - C

Temperature	-	1200° F
Pressure	-	30 psia
Mass Flow Rate	-	183,300 lb./hr.
Enthalpy	-	32.40×10^6 Btu/hr.

FLOW SHEET

Feed Water Stream - W

Temperature	-	100°F
Pressure	-	680 psia
Mass Flow Rate	-	136,800 lb/hr.
Enthalpy	-	7.80×10^6 Btu/hr.

Feed Water Stream - $w_1 = w_2 = w_3 = w_4$

Temperature	-	100°F
Pressure	-	680 psia
Mass Flow Rate	-	34,200 lb./hr.
Enthalpy	-	1.95×10^6 Btu/hr.

Steam from Reactor - $S_1 = S_2 = S_3 = S_4$

Temperature	-	500°F
Pressure	-	680 psia
Mass Flow Rate	-	34,200 lb./hr.
Enthalpy	-	172.2×10^6 Btu/hr.

Combined Steam - S

Temperature	-	500°F
Pressure	-	680 psia.
Mass Flow Rate	-	136,800 lb./hr.
Enthalpy	-	688.8×10^6 Btu/hr.

FLOW SHEET

Control Point 1

Control and records rate of flow of fresh feed gas and temperature and pressure of stream.

Control Point 2

~~Con~~ Record temperature and pressure of recycle stream

Control Point 3

Samples of regenerated catalyst will be taken here. The rate of flow of regenerated catalyst is recorded.

Control Point 4

The pressure and rate of flow of air to the regenerator is recorded here.

Control Point 1 $S_1 = S_2 = S_3 = S_4$

Samples of the catalyst from the reactor are tested for fouling and depending upon the results the amount of catalyst to regeneration and air for regeneration is controlled.

FLOW SHEET

Control Point - $5_1 = 5_2 = 5_3 = 5_4$

The temperature, pressure, and mass rate to each reactor is recorded here.

Control Point - $6_1 = 6_2 = 6_3 = 6_4$

The mass rate of water to the steam tubes in the reactor is recorded and controlled at this point.

Control Point - $7_1 = 7_2 = 7_3 = 7_4$

The temperature and mass rate of the product from each reactor is recorded here.

Control Point 8

The temperature and mass flow rate of stack gases from the regenerator is recorded at this point

Control Point - $9_1 = 9_2 = 9_3 = 9_4$

The temperature at various points through the reactors is recorded and used to control the rate of fresh feed and the steam pressure.

FLOW SHEET

Control Point - 10

The temperature through the regenerator is recorded.

Control Point - $11_1 = 11_2 = 11_3 = 11_4$

The mass rate of flow and the pressure of the steam from each set of reactor cooling tubes is recorded and the pressure controlled

Control Point - $12_1 = 12_2 = 12_3 = 12_4$

The pressure existing in the reactor body is recorded here.

Control Point - 14

The mass flow rate of Fresh Catalyst is recorded here.

Bibliography

1. Keith, P. C. Oil and Gas J. 45, No 6, 102 (1946)
   Chem. Eng 53, 101 (1946)
2. Fischer, F. and Tropsch, H. Brennstaff-Chem 4, 276 (1923)
3. Fischer, F. and Tropsch, H. Brennstaff-Chem 7, 97 (1926)
   Ber. 59, 830, 832, 923 (1926)
   Fischer, F. and Meyer K., Brennstaff Chem 12, 225 (1931)
   Fischer, F., Pichler, H., and Reder, R., Brennstaff Chemic 13, 421 (1932)
   Fischer, F. Roelen, O., and Fiesst, W., Brennstaff Chemic, 13, 461 (1932)
   Fischer, F., and Koch, H., Brennstaff Chem. 13, 428 (1932)
4. Fischer, F., and Dilthey, P., Brennstaff Chem.9, 24 (1928)
5. Richler, H., and Merkel, H., U.S. Bureau of Mines Special Report (1947)
6. Emmet, P. H. et. al. Proc. of Am. Chem. Soc. April, 1947
7. Storch, H. H., Chem. Eng. Prog. 44, No. 6, 469 (1948)
8. Storch, H. H., Anderson, R. B., Hofer, L. J. E., Hawk, C. O., and Golumbic, N., Bureau of Mines Technical Paper 709 (1949)
9. Ryan, P., Oil and Gas J. 43, No. 47, 264 (1945)
10. Roberts, J. G., and Phinney, J. A., Oil and Gas J. 45, 72, 139 (1947)
11. Anon, Oil and Gas J. 45, 58 (1946)
12. Eliot, E., Goddin, A., Pace, R., Chem. Eng. Prog. 44, No 8, 533 (1948)

Product Separation - IV

Urbana, Illinois
May 31, 1950

Dr. H. F. Johnstone
Division of Chemical Engineering
University of Illinois
Urbana, Illinois

Dear Sir:

The following report includes all design specifications and costs for the section of a Fischer-Tropsch synthesis plant required for the separation of products.

Any further information desired will gladly be furnished upon request.

Very truly yours,

Arnold P. Schueler

Arnold P. Schueler
Product Separation Design Group

APS/RS

SEPARATION OF PRODUCTS FROM FISCHER-TROPSCH SYNTHESIS

Chemical Engineering Division
University of Illinois
Urbana, Illinois

Plant Design Group

Caskey
Chiang
Houvouras
Koeller
Schueler
Junk

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SUMMARY

Separation of Products from Fischer-Tropsch Synthesis

At the request of Dr. H. F. Johnstone, head of Chemical Engineering Division, the plant required for the separation of products from a Fischer-Tropsch synthesis using natural gas to produce 5,000 barrels of gasoline per day was designed to determine the pre-pilot plant first cost approximation.

Reactor gaseous products at a rate of 2,346,500 pounds per hour at 300 psia and 600°F. (equivalent to approximately 5,000 barrels of gasoline per day) can be separated into four main streams, water and water soluble chemicals at 1,486,344 pounds per hour, oil and oil soluble chemicals at pounds per hour, light paraffin hydrocarbon gases at pounds per hour, and recycle gases for the reactor at pounds per hour. These separations can be accomplished at a total investment of $ and operating expenses of $ per year, based on 8,760 hours of operation per year.

The reactor gaseous products are partially condensed in a direct contact spray condenser (see Flow Diagram, pg. 6 and Table I). The condensed products are separated into two layers, oil and oil soluble chemicals and water and water soluble chemicals in tank gravity separators. The oil layer is sent to the refining process without further purification. Part of the water layer is recycled through a tube and shell heat exchanger for use as spray condenser coolant.

The non-condensed gases pass through a water scrubber in which the remainder of the water-soluble chemicals are removed and combined with the unrecycled portion of the water layer from

the condenser. The combined stream is then sent to the chemicals recovery and purification plant.

The gases from the water scrubber pass through a straw-oil absorber where light hydrocarbon gases are removed. The resulting gas stream is then recycled to the reactor feed after suitable bleed to prevent accumulation of undesired inerts. The light hydrocarbon gases are stripped from the straw-oil and proceed to the alkylation and polymerization plants.

[illegible] the straw-oil absorber & stripper [illegible] able to complete his work.

[illegible] reactor group just recently changed reactor [illegible] stream to a small degree. These changes are [illegible] included in the design because of lack of time [illegible] the comprehensive design of the following units should [illegible] to enable present equipment to handle the [illegible] change.

INTRODUCTION

At the request of the Chemical Engineering Division a Fischer-Tropsch synthesis plant to produce 5,000 barrels of gasoline per day from natural gas was designed to determine first cost approximations. This report includes only the design of the section of the plant required for the separation of the products from the main reactor.

Little information and data on these steps in the production of gasoline by Fischer-Tropsch synthesis were available in the literature. The brief descriptions of the war-time German plants indicated that the semi-batch processes employed were hardly applicable to an efficient continuous process plant. In addition the proposed plant was designed for a maximum yield of gasoline whereas, the German processes were not designed on this basis. Many of the physical data required for accurate design were also found to be lacking.

This report contains a description of this section of the overall process, a process flowsheet showing the design flows of heat and material, the estimated physical properties used and the methods used in their estimation, design calculations and specifications, and a first estimate of total investment and operating costs.

DESCRIPTION OF PROCESS

The reactor gaseous products at 600°F. and 300 psia at a rate of 2,346,500 pounds per hour (for distribution see Table II) enter counter-current spray condenser, C, where partial condensation occurs when the gases are cooled to 200°F. (see pages 6 and 7). The condensed stream, including the coolant sprayed into the condenser, then passes to a liquid gravity separation tank, LS, where two layers are formed, an oil and oil-soluble chemicals layer and a water and water-soluble chemicals layer. The oil layer, 5b, passes directly to the refining process to obtain gasoline.

Part of the water layer, 5a, goes directly to the chemicals recovery plant. The remaining portion, 5d, is recycled through a water-cooled tube and shell heat exchanger, IC, where it is cooled to 120°F. The cooled stream is then sprayed into the condenser.

The uncondensed vapors, 5c, pass from the condenser to a plate column water-scrubber, GS, where the products are cooled to 85°F. and the remaining water-soluble chemicals are removed. The bottom stream, 10, is combined with the unrecycled portion of the water layer from the separator.

The vapors from the water scrubber,7, pass through straw-oil absorber, A, where the light hydrocarbon gases are absorbed. The discharging gas stream, 8, is bled, 3, continuously to reduce accumulation of inerts and the remaining portion, 4, is compressed to 325 psia and recycled to the reactor.

The absorber bottoms are stripped of light hydrocarbons in steam stripper, S, from which the bottoms, 11, are recycled to the absorber. The stripped gases, 9, pass to the alkylation and polymerization plants.

Quantities, temperatures and pressures of all streams and equipment specifications are given on the flow sheet and in Table I.

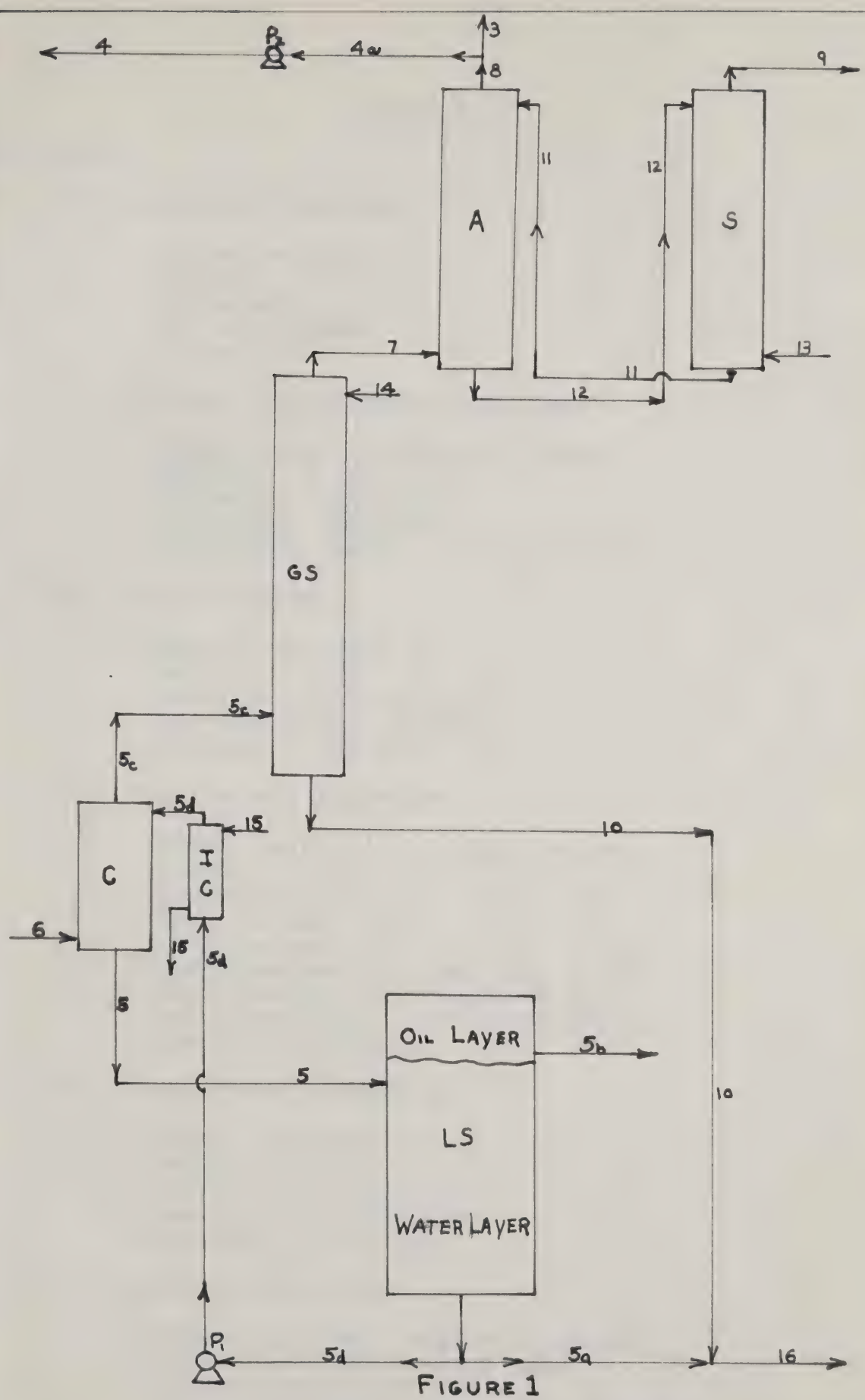


FIGURE 1
SIMPLIFIED FLOW DIAGRAM
PRODUCT SEPARATION

TABLE I

Equipment

A - Straw-Oil Absorbers

Number required:
Height:
Diameter:
No. of Plates:
Pressure:

C - Direct Contact Spray Condensers

Number required: 4 plus 1 spare
Height: 17 ft.
Diameter: 12.5 ft.
Pressure: 300 psia
Total Heat Load: 411.18 x 10^6 Btu/hr.

GS - Water Scrubbers

Number required: 3
Height: 76 ft.
Diameter: 11.25 ft. I.D.
No. of plates: 38
Pressure: 300 psia

IC - Recycle Heat Exchanger

Number required: 4 plus 1 spare
Length: 16 ft.
Diameter: 7 ft.
Shell passes: 6
Tube passes: 8
Tube diameter: 1 inch O.D. 18 BWG
Heat transfer area: 12,500 $ft.^2$
Total heat load: 411 x 10^6 Btu/hr.
Pressure: 350 psia

LS - Liquid Tank Separators

Number required: 12
Height: 50 ft.
Diameter: 25 ft.
Capacity: 183,600 gals.
Pressure: 300 psia

P_1 - Liquid Recycle Pump

Number required: 4 plus 2 spares
Type: Heavy duty centrifugal
Driving motor: 125 Hp
Duty: 2570 gpm from 300 to 360 psia

TABLE I - Cont.

P_2 - Vapor Recycle Compressor

S - Steam Stripper

Number required:
Height:
Diameter:
No. of plates:
Pressure:

TABLE II

Stream Compositions and Flow Rates
(All enthalpies based on 32°F.)

Stream (6)

Component	Moles/Hr.	lbs./hr.
CO_2	18,520	815,000
CO	16,000	448,000
H_2	32,620	65,300
N_2	6,990	195,800
H_2O	6,975	125,600
C_1-C_2*	3,000	54,000
C_3-C_4*	1,298	64,900
C_5-C_{10}*	1,340	142,100
C_{11}-C_{18}*	197	402,000
Alcohols	245	12,750
Aldehydes	49	2,550
Ketones	44	2,640
Acids	103	6,590
Oil Soluble Chemicals	80	9,270
	87,461 moles/hr	2,345,500 lbs/hr

Temperature - 600°F.
Pressure - 300 psia
Enthalpy - 616,305,000 Btu/hr.

Stream (5c)

Component	Moles/hr.	lbs./hr.
CO_2	18,520	815,000
CO	16,000	448,000
H_2	32,620	65,300
N_2	6,990	195,800
H_2O	3,150	56,700
C_1-C_2*	2,986.4	53,700
C_3-C_4*	1,232.4	61,600
C_5-C_{10}*	824.0	87,300
C_{11}-C_{18}*	14.0	2,855
Alcohols	239.1	12,450
Aldehydes	47.8	2,490
Ketones	43.0	2,580
Acids	100.9	6,460
Oil Soluble Chemicals	48.7	5,640
	82,816.3	1,815,875

Temperature - 200°F.
Pressure - 300 psia
Enthalpy - 193,497,000 Btu/hr.

* Include saturated and unsaturated hydrocarbons.

Stream (5a)

Component	Moles/hr.	lbs./hr.
H_2O	3825.0	68,900.0
Alcohols	5.9	307.0
Aldehydes	1.2	62.3
Ketones	1.0	60.0
Acids	2.1	134.5
	3835.2	69,463.8

Wgt. per cent chemicals 0.79%
Temperature: 200°F.
Pressure: 300 psia
Enthalpy: 11,561,070 Btu/hr.

Stream (5b)

Component	Moles/hr.	lbs./hr.
C_1-C_2	13.6	244.6
C_3-C_4	65.6	3,280.0
C_5-C_{10}	516.0	54,700.0
C_{11}-C_{18}	183.0	37,350.0
Oil Soluble Chemicals	31.3	3,630.0
	809.5	99,204.6

Temperature: 200°F.
Pressure: 300 psia
Enthalpy: 59,850 Btu/hr.

Stream (5d)

Same composition as (5a) 5.14×10^6 lbs./hr.

Enthalpy entering heat exchanger = 863,000,000 Btu/hr. 200°F. 330 psia
" leaving " " = 452,000,000 Btu/hr. 120°F. 325 psia

Stream (5)

(5) = (5a) + (5b) + (5d) = 5,308,669 lbs/hr.

Temperature: 200°F.
Pressure: 300 psia
Enthalpy: 874.62×10^6 Btu/hr.

Stream (15)

Water 4.11×10^6 lbs/hr. = 494,000 gph
Enthalpy entering heat exchanger = 197×10^6 Btu/hr. 80°F.
" leaving " " = 608×10^6 Btu/hr. 180°F.

Stream (10)

Composition	Moles/hr.	lbs./hr.	Wgt. %
H_2O		1,392,900	
Alcohols	239	12,450	0.880
Aldehydes	48	2,490	0.176
Ketones	43	2,580	0.182
Acids	101	6,460	0.456
		1,416,880	

Temperature: 190°F.
Pressure: 300 psia
Enthalpy: 211,347,000 Btu/hr

Wgt. % Chemicals = 1.694

Stream (14)

Water = 1,339,000 lbs./hr. = 160,800 gph
Enthalpy = 64,250,000 Btu/hr.
Temperature = 80°F.

Stream (7)

Component	Moles/hr.	lbs./hr.
CO_2	18,520	815,000
CO	16,000	448,000
H_2	32,620	65,300
N_2	6,990	195,800
H_2O	156	2,800
C_1-C_2	2,986	53,700
C_3-C_4	1,232	61,600
C_5-C_{10}	824	87,300
C_{11}-C_{18}	14	2,855
Oil Soluble Chemicals	49	5,685
	78,391	1,736,995

Temperature: 85°F.
Pressure: 290 psia
Enthalpy: 46,400,000 Btu/hr.

Stream (8)

Stream (3)

Same composition as Stream (8) lbs./hr.

Temperature	°F.
Pressure	psia
Enthalpy	Btu/hr.

Stream (12)

Component	Moles/hr.	lbs/hr.

Temperature	°F.
Pressure	psia
Enthalpy	Btu/hr.

Stream (11)

Component	Moles/hr.	lbs./hr.

Temperature	°F.
Pressure	psia
Enthalpy	Btu/hr.

Stream (9)

Component	Moles/hr.	lbs./hr.

Stream (13)

Stream (4a)

Composition same as Stream (8): lb./hr.

Temperature	°F.
Pressure	psia
Enthalpy	Btu/hr.

Stream (4b)

Overall Energy Balance

Basis: 1 hour

	Btu/hr.	Per cent
Input:		
Reactor Products	616,305,000	
Cooling Water		
Recycle Heat Exchangers	197,000,000	
Water Scrubbers		
Gas Compressors		
psig Steam to Strippers		
Output:		
Oil and Oil-Soluble Chemicals	59,850+	
Water and Water-Soluble Chemicals (10) and (5a)		
Light Hydrocarbon Gases (9)		
Bleed (3)		
Reactor Recycle		
Cooling Water		
Recycle Heat Exchangers	608,000,000	
Gas Compressors		

Overall Material Balance

Basis: 1 hour

Input:	Pounds/hr.
Reactor Products	2,346,500
Cooling Water	
Recycle Heat Exchangers	4,110,000
Water Scrubbers	
Gas Compressors	
Steam to Strippers	
Make-up Oil	

Output:	
Oil and Oil Soluble Chemicals	99,205
Water and Water-Soluble Chemicals (10) and (5a)	
Light Hydrocarbon Gases	
Bleed	
Reactor Recycle	
Cooling Water	
Recycle Heat Exchangers	4,110,000
Gas Compressors	

PROCESS EQUIPMENT SPECIFICATIONS AND COSTS

Vessels	No. Required	Diameter Ft.	Hgt. Ft.	Operating Press. Psia	Unit List $	Instal-lation Cost $	Total Unit Cost $	Total Cost $
A - Straw-Oil Absorber (Plate Column)								
C - Direct Contact Spray Condenser	5	12.5	17	300	11,330	11,330	22,660	113,300
GS- Water Scrubber (Plate Column)	3	11.25	76	300	103,900	44,100	148,000	444,000
LS - Liquid Tank Separators	12	25	50	300	90,000	9,000	99,000	1,188,000
S - Stripper								

Heat Exchangers	No. Required	Heat Trans Area - Ft.^{2}	Operating Press. Psia	Unit List	Instal-lation Cost	Total Unit Cost	Total Cost
IC - Recycle Exchanger	5	12,500	350	39,600	17,000	56,600	283,000

Pumps	No. Required	gpm	Hp	Total Head Psia	Unit List	Instal-lation Cost	Total Unit Cost	Total Cost
P_1 - Liquid Recycle Pump	6	2570	125	60	4,130	1,770	5,900	35,400

Compressors	No. Required	No. of Stages	Lbs./hr.	Hp	Total Head	Unit List	Instal-lation Cost	Total Unit Cost	Total Cost
P_2 - Vapor Recycle Compressor									

PROCESS EQUIPMENT SPECIFICATIONS AND COSTS
(Cont.)

Instruments	No. Required	Unit List	Access-ories	Installation	Total Unit	Total Cost
Orifice Meter Recording and Controlling	5	600	240	240	1080	5,400
Pneumatic Control System	37	300	45	75	420	15,540
Pneumatic Transmission Components	36	175	16	44	235	8,460
Pressure Recorder	5	150	45	60	255	1,275
Temperature Recorder and Controller	10	675	270	338	1283	12,830
Orifice Meter Recording	7	500	200	200	900	6,300
Automatic Timing Device	4	1000		500	1500	4,500
Flow Recorder and Controller	3	160	152	80	392	1,176
Thermocouples	3	120	-	-	120	360
Pneumatic Control Systems	3	300	45	135	480	1,440
3-Point Temp. Recorder	2	800	-	400	1200	2,400

TOTAL INVESTMENT COST

A. Basic Process Equipment Cost
(Including Installation)

1. Vessels
2. Heat Exchangers
3. Pumps
4. Compressors

B. Auxiliary Equipment Cost

1. Piping
   a. Equipment
   b. Installation

2. Instrumentation
   a. Equipment
   b. Installation

3. Electric Wiring
   a. Equipment
   b. Installation

4. Insulation, Fireproofing Etc.
   a. Equipment
   b. Installation

OPERATING COSTS

Basis:- 1 year (8760 hrs.)

A. Steam

B. Cooling Water

C. Straw-Oil Make-up

D. Electricity

E. Labor

F. Depreciation

G. Taxes, Maintenance,
Insurance and Patent Rights

ECONOMIC CONSIDERATIONS

The first important consideration in this design was the type of condenser to be used for the reactor products. The three types considered were tube and shell, venturi-spray, and direct contact spray condensers. The first type was eliminated when the large quantities of gases to be handled, of which 85 volume % were non-condensible, were considered. The extremely low vapor coefficients of heat transfer under these conditions would necessitate unreasonably large heat transfer areas.

The venturi-spray condenser also proved inadequate because of the large quantities of coolant required to remove the heat load. Consequently, a direct-contact low velocity spray condenser was selected.

The first design on the spray condenser included the use of a surface cooler to cool a recycled portion of the water and water-soluble chemicals used as coolant in the condenser. It was then considered that since the recycle stream was practically pure water (0.266 mole % chemicals corresponding to a 200°F. condenser outlet temperature) it would not be economically sound to use a recycle cooler unless the temperature rise of the cooling water was more than the rise in temperature of the recycle stream as it passed through the condenser. The elimination of the recycle cooler, however would require that the large quantities of coolant water be sprayed directly into the condenser and consequently the water-soluble chemicals stream would be made even more dilute (0.000548 mole % chemicals) and the subsequent recovery of the chemicals would become more costly. If this dilute stream were discarded as waste, a total of $41,000 per year would be lost. (Prices of chemicals based on that value for April 1950 and assuming only 10% of these prices as profit.)

From the concentration figures of the water-soluble chemicals stream, it was obvious that the purification difficulties and costs for the more dilute stream would be increased beyond reason. Therefore, if the recycle cooler were eliminated, this dilute stream would have to be discarded. On the other hand, if the chemicals were to be recovered from this stream, the recycle cooler would be required to keep the separation costs as low as possible.

The compositions of the water and water-soluble chemicals stream leaving the water scrubber was 0.603 mole % chemicals corresponding to a condenser outlet temperature of 200°F. These chemicals must be recovered for overall economic reasons. If the recycle cooler was not eliminated and the water and chemicals stream from the condenser were combined with that from the scrubber, the resulting stream would contain 0.586 mole % chemicals. Obviously from these figures, the increased difficulty and cost of purification of the combined stream above that of the stream from the scrubber was negligible. Consequently, it was concluded that it would not pay to discard the condenser chemical stream and it followed that the recycle cooler was required.

The next consideration was the selection of the outlet temperature from the spray condenser. A lower limit of 120°F. was set because of the limitations imposed by the 80°F. cooling water available for the recycle cooler. Calculations on the partial condensation obtained, the cooling water required and the recycle cooler costs were made at various condenser outlet temperatures from 120°F. to 200°F. Overall costs decreased with increasing outlet temperature.

However, similar calculations on the costs of the water scrubber indicated that lower condenser outlet temperatures favored lower costs. Little or no economic considerations from this point of view were necessary for the oil absorber since nearly the same quantities were handled at all the different condenser outlet temperatures. Consequently, an economic balance between water scrubber and spray condenser and intercooler costs was made to select the proper condenser outlet temperature. (See Figure 2 Pg.22) A broad minimum was indicated and a condenser outlet temperature of 200°F. was selected as economically sound.

The economic considerations made in the design of each individual unit are included in the design calculations.

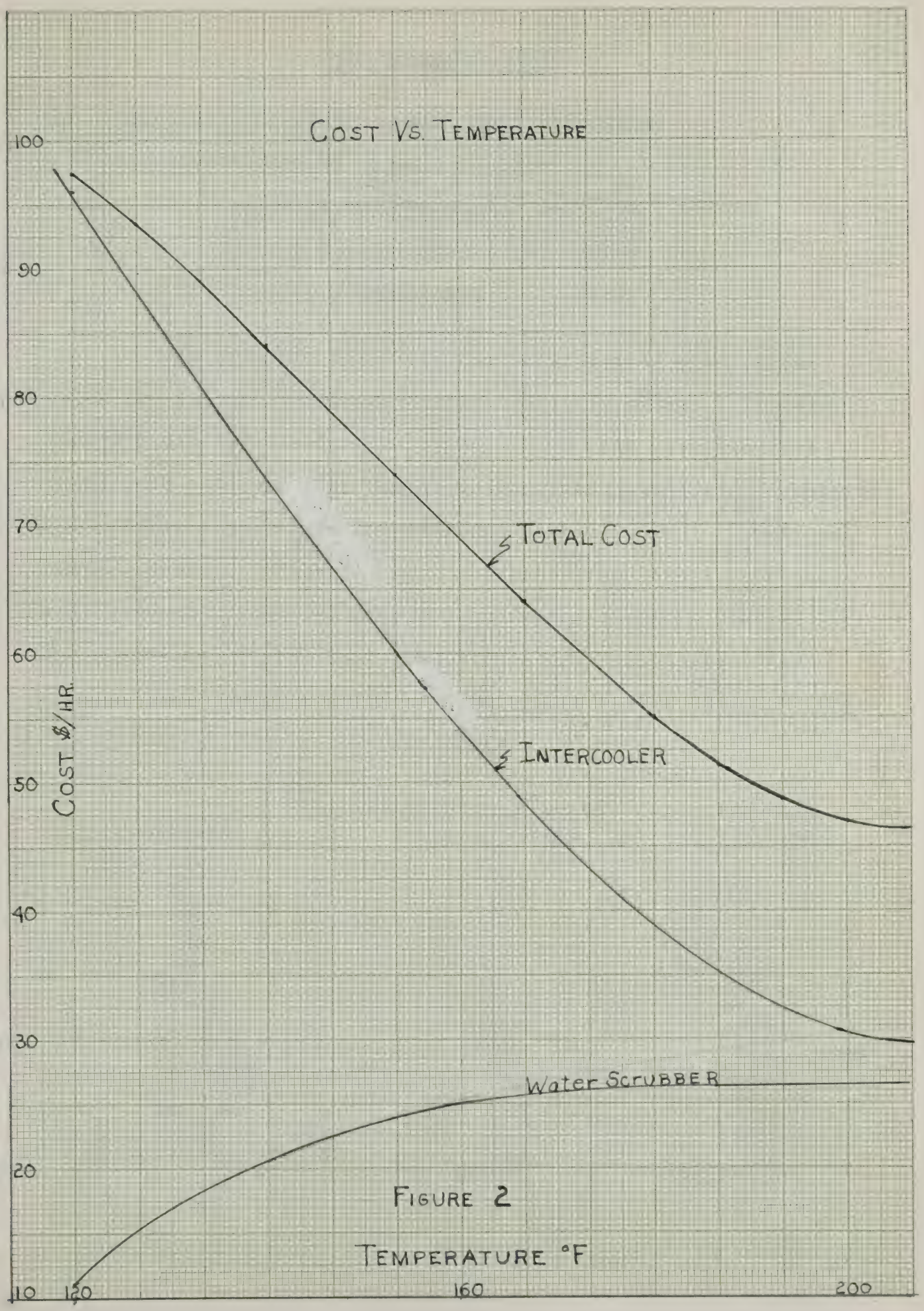


FIGURE 2

NATIONAL
No. 5780-20
20 SQUARES TO THE INCH
Made in U. S. A.

DESIGN CALCULATIONS

I. Condensers, Separators, and Recycle Coolers

A. Direct Contact Spray Condenser

1. Calculation of Amounts Condensed

Vapor Inlet Temperature: 600°F.

Outlet Temperature: 200°F.

In order to calculate the vapor-liquid equilibrium split for this multicomponent system at this outlet temperature, vapor-liquid equilibrium constants at 300 psia were required. These were obtained from direct data or estimated from fugacities and reduced temperature and pressure plots.[1, 2, 3, 4] Critical temperatures and pressures were obtained from actual data and approximated by the use of petroleum characteristic factors[5, 6, 7, 8]. These data are listed in Table III Page 41. Equilibrium flash calculations showed a split of 5.3% liquid and 94.7% vapor with the following compositions and rates:

Component	Moles/hr. Liquid	Moles/hr. Vapor
CO_2	0	18,520.0
CO	0	16,000.0
H_2	0	32,620.0
N_2	0	6,990.0
H_2O	3,825.0	3,150.0
C_1-C_2	13.6	2,986.4
C_3-C_4	65.6	1,232.4
C_5-C_{10}	516.0	824.0
C_{11}-C_{18}	183.0	14.0
Alcohols	5.9	239.1
Aldehydes	1.2	47.8
Ketones	1.0	43.0
Acids	2.1	100.9
Oil Soluble Chemicals	31.3	48.7
	4644.7	82,876.3

2. Enthalpy Balance

In order to calculate enthalpies vapor specific heat data for various temperatures at 300 psia were obtained from actual data or estimated from reduced temperature and pressure plots[5, 6, 8, 9]. Latent heats of vaporization were similarly obtained[5, 6, 10]. These data are tabulated in Table III Page 41. Base Temperature for all enthalpies is 32°F.

a. Enthalpy of Vapor Stream Entering Condenser - H_6

$$H = (\text{moles/hr.})(\text{Btu/lb.-mole}^{\circ}\text{F.})(600 - 32) + (\text{moles/hr.})(\text{Btu/lb.-mole})$$

Component			Hi Btu/hr.
CO_2	(18,520)(11.1)(600-32)	=	116,900,000
CO	(16,000)(7.25)(568)	=	65,900,000
H_2	(32,620)(6.95)(568)	=	129,000,000
N_2	(6,990)(7.15)(568)	=	28,400,000
H_2O	(6,975)(1202.8)(18)	=	151,000,000
C_1-C_2	(3,000)(13.2)(568)	=	22,500,000
C_3-C_4	(1,298) [5250 + 27.1 (568)]	=	26,800,000
C_5-C_{10}	(1,340) [5720 + 56 (568)]	=	50,200,000
C_{11}-C_{18}	(197) [10,000 + 105 (568)]	=	13,700,000
Alcohols	245 [7,500 + 24 (568)]	=	5,170,000
Aldehydes	49 [7,500 + 19 (568)]	=	896,000
Ketones	44 [7,500 + 24 (568)]	=	929,000
Acids	103 [9,000 + 21 (568)]	=	2,160,000
Oil Soluble Chemicals	80 [10,000 + 43 (568)]	=	2,750,000
		H_6 =	616,305,000 Btu/hr.

b. Calculation of Enthalpy of Vapor Stream Leaving Condenser - H_{5c} - at 200°F.

Component			Hi Btu/hr.
CO_2	(18,520)(11)(200 - 32)	=	34,300,000
CO	(16,000)(7.23)(168)	=	19,420,000
H_2	(32,620)(6.9)(168)	=	37,850,000
N_2	(6,990)(7.2)(168)	=	8,400,000
H_2O	(3,150) [(809)(18) + (8.6)(168)]	=	50,500,000
C_1-C_2	(2,986.4) [30000 + (11.3)(168)]	=	14,600,000

(Continued on next page)

Component - Continued			Hi Btu/hr.
C_3-C_4	(1232.4)[5250 + (24.5)(168)]	=	11,540,000
C_5-C_{10}	(824)[5720 + (46)(168)]	=	11,080,000
C_{11}-C_{18}	(14)[10,000 + 86 (168)]	=	343,000
Alcohols	(239.1)[7,500 + 20 (168)]	=	2,600,000
Aldehydes	(47.8)[7,500 + 16.6 (168)]	=	492,000
Ketones	(43.0)[7,500 + 18.6 (168)]	=	456,000
Acids	(100.9)[9,000 + 14.5 (168)]	=	1,151,000
Oil Soluble Chemicals	(48.7)[10,000 + 34 (168)]	=	765,000
		H_{5c} =	193,497,000 Btu/hr.

c. Calculation of Enthalpy of Oil and Oil Soluble Chemicals Stream H_{5b}

Component		Hi Btu/hr.
C_1-C_2	(13.6)(0.55)(200 - 32)	1,250
C_3-C_4	(65.6)(0.43)(168)	4,750
C_5-C_{10}	(516.0)(0.43)(168)	37,200
C_{11}-C_{18}	(183.0)(0.44)(168)	13,500
Oil Soluble Chemicals	(31.3)(0.60)(168)	3,150
	H_{5b} =	59,850 Btu/hr.

d. Calculation of Enthalpy of Unrecycled Water and Water Soluble Chemicals Stream - H_{5a}

This stream will remove the water and water soluble chemicals formed in the reaction less that portion which goes out of condenser as vapor.

Component		Hi Btu/hr.
H_2O	(3825)(18)(168)	11,560,000
Alcohols	(5.9)(0.668)(168)	664
Aldehydes	(1.2)(0.67)(168)	135
Ketones	(1.0)(0.54)(168)	91
Acids	(2.1)(0.51)(168)	180
	H_{5a} =	11,561,070 Btu/hr.

e. Calculation of Enthalpy of Recycle Stream - H_{5d}

Condenser Heat Load = $H_6 - H_{5a} - H_{5b} - H_{5c}$ =

411,180,000 Btu/hr.

The composition of the recycle stream is the same as that of stream 5_a.

$$\text{Mole \% Chemicals} = \frac{(10.2)(100)}{3825 + 10.2} = 0.266\%$$

This is essentially pure water and therefore it is treated as such in the following calculations.

Condenser Inlet Temperature of Recycle Stream 120°F.

Condenser Outlet " " " " 200°F.

Recycle Required = $\frac{411,180,000}{200 - 120}$ = 5,140,000 lbs./hr = 616,000 gph

H_{5d_1} = Enthalpy of Recycle Stream Entering Condenser =

$(5.14 \times 10^6)(1)(120 - 32) = 452,000,000$ Btu/hr.

H_{5d_2} = Enthalpy of Recycle Stream Leaving Condenser =

$(5.14 \times 10^6)(1)(200 - 32) = 863,000,000$ Btu/hr.

f. Calculation of Enthalpy of Stream 5 Leaving Condenser H_5

$H_5 = H_{5a} + H_{5b} + H_{5d_2} = 874,600,000$ Btu/hr.

g. Complete Heat Balance on Condenser

Input = $H_6 + H_{5d_1} = 1068.3 \times 10^6$ Btu/hr.

Output = $H_{5c} + H_5 = 1068.1 \times 10^6$ Btu/hr.

3. Material Balance

Input = $w_6 + w_{5d}$ = 2,346,500 + 5,140,000 = 7,486,500 lbs./hr.

Output = $w_{5c} + w_5$ = 1,815,875 + 5,308,669 = 7,124,544 lbs./hr.

4. Dimensional Design

For lack of a better design method the empirical equation as given in the 1950 A.I.Ch.E. contest problem was employed. It was realized that this is probably inaccurate but it was felt that it was more suitable than the design methods for steam spray condensers, the only other methods available. According to this equation

$$U_{max} = \frac{0.136\sqrt{\rho_L - \rho_G}}{\rho_G}$$

where $U_{max.}$ = maximum vapor velocity - ft./sec.

ρ_L = density of liquid - lbs./ft.^{3}

ρ_G = density of vapor - lbs./ft.^{3}

a. Height

The height for use with this equation was arbitrarily set at 17 feet.

b. Diameter

$$U_{max.} = \frac{0.136\sqrt{62.4 - 0.588}}{0.588} = 1.82 \text{ ft./sec.}$$

Entering gas rate = 87,461 moles/hr. = 3.31×10^6 ft.^{3}/hr.

Design for four condensers plus one spare

Cross-sectional area of flow = $3.31 \times 10^6/(3600)(4)(1.82)$

= 126 ft.^{2}

$$\text{Diameter} = \sqrt{\frac{126 \times 4}{\pi}} = 12.65 \cong 12.5 \text{ ft.}$$

c. Thickness of Shell

Safety factor = 5

Yield strength = 60,000 psi

Working Pressure = 300 psi

t = thickness r = radius

$$t = \left[\sqrt{\frac{\frac{60,000}{5} + 300}{\frac{60,000}{5} - 300}} - 1\right] \times r = 0.025r$$

Thickness of shell = $(0.025)(\underline{12}.5)(12)/(2) = 1.87$ inches

d. Number of Spray Nozzles[15]

Nozzle type and size: 0.5 inch Orifice. Solid Cone Nozzles 84° Angle

Pressure Drop - 25 psi

Flow Rate = 27.5 gpm

Number of nozzles = 2570 gpm/27.5 = 94 nozzles/condenser

The interior construction and nozzle arrangement is to be left to the discretion of manufacturer.

e. Type of Material

Ordinary steel may be used in fabrication since very little corrosive chemicals are present.

5. Cost Calculations

a. Vessel Cost = ($303/ft. length)(17 ft.)(2.2 press. factor)
= $11,330 per condenser

b. Construction Costs

According Vilbrandt, "Chemical Engineering Plant Design" page 421, an estimation of total cost including construction can be obtained by dividing initial cost by 0.7. Since little was known about the internal construction a conservative construction cost value was obtained by dividing by 0.5.

Construction Cost = $11,330 per condenser

c. Total Cost = ($22,660)(5) = $113,300

6. Pressure Drop

Assume negligible pressure drop across the condensers.

B. Recycle Cooler

1. Heat Balance

Cooling Water Inlet Temperature - 80°F.

Cooling Water Outlet Temperature - 180°F.

Recycle Inlet Temperature - 200°F.

Recycle Outlet Temperature - 120°F.

Enthalpy of Entering Recycle Stream = H_{5d_2} = 863,000,000 Btu/hr.

" " Exit " " = H_{5d} = 452,000,000 Btu/hr.

Heat Load = Q = 863,000,000 - 452,000,000 = 411,000,000 Btu/hr.

Cooling Water Required = $\frac{411,000,000}{(180 - 80)}$ = 4,110,000 lbs./hr.

= 494,000 g.p.h.

Enthalpy of Inlet Water = (4,110,000)(80 - 32) =

197,000,000 Btu/hr.

Enthalpy of Outlet Water = (4,110,000)(180 - 32) =

608,000,000 Btu/hr.

Input = $197 \times 10^6 + 863 \times 10^6 = 1060 \times 10^6$ Btu/hr.

Output = $608 \times 10^6 + 452 \times 10^6 = 1060 \times 10^6$ Btu/hr.

2. Material Balance

Input = 4.11×10^6 lbs./hr. + 5.14×10^6 lbs./hr.

= 9.25×10^6 lbs./hr.

Output = $4.11 \times 10^6 + 5.14 \times 10^6 = 9.25 \times 10^6$ lbs./hr.

3. Calculation of Heat Transfer Area Required

An economic balance on heat exchanger cost, water costs, and pressure drop costs indicated a tube side velocity of three ft./sec., an overall heat transfer coefficient of 300 Btu/hr. ft.^{2} °F. an outlet temperature on tube side of 180°F. and 1-inch O.D. tubes

for minimum costs. (Recycle on tube side, water on shell side.) This balance also indicated the use of six shell passes and an even number of tube passes[11].

$U = 300$ Btu/hr. ft.^{2} ^oF.

Ln Mean $\Delta t = \frac{40 - 20}{\ln 2} = 28.8^oF$.

Correction factor for temperature difference for multipass exchangers[11] = 0.95

Heat Transfer area required $= \frac{411 \times 10^6}{(300)(0.95)(28.8)} = 50{,}000$ ft.^{2}

4. Material of Construction

The recycle stream consists of extremely dilute organic acids mainly acetic acid. Using the data in Vilbrandt, page 287 on the corrosiveness of chemicals, mild carbon steel was selected as suitable construction material.

5. Cost Calculations

It was decided to use one exchanger with each condenser, or a total of five, in order to provide more flexible operation. Only four of the five will bein active service at a time.

Heat transfer area per recycle cooler $= \frac{50{,}000}{4} = 12{,}500$ ft.^{2}

Cost data for tube and shell floating head exchanger from Vilbrandt page 468.

a. Cost per exchanger = (\$2.00/ft.^{2})(12,500)ft.^{2}(1.2)(1.32[14])
press. factor, price index factor

= \$39,600

b. Installation Cost = \$17,000 per exchanger

This was obtained using the estimate proposed by Vilbrandt page 421.

c. Total Equipment Cost = (5)(39,600 + 17,000) = \$283,000

d. Cooling Water Costs

Cooling tower and spray ponds were found impractical in high humidity of Louisiana. Filtered river water at 4¢ per 1000 gals. will be used.

Cooling water costs = $(0.494 \times 10^6 \text{ gph})(\$0.04 \text{ per } 1000 \text{ gals})$

$= \$19.76/\text{hr.}$

6. Dimensional Design

a. Number of tubes per exchanger

Recycle Rate = $(5.14 \times 10^6 \text{ lb./hr.})\left(\frac{1}{4 \text{ exchangers}}\right)\left(\frac{1}{62.2 \text{ lb/ft}^3}\right)$

$= 2.06 \times 10^4 \text{ ft.}^3/\text{hr.}$

Total Cross-sectional Area of Flow =

$(2.06 \times 10^4 \text{ ft.}^3/\text{hr})\left(\frac{1}{3 \text{ ft./sec.}}\right)\left(\frac{1}{3600 \text{ sec./hr.}}\right) =$

1.91 ft.^2

Using a safety factor of 5 and yield stress of 60,000 psi and working pressure of 300 psi tube thickness = 0.0125 inch. 1-inch 18 BWG Steel Tubing satisfies this pressure requirement.

No. of Tubes = $(1.91 \text{ ft.}^2)\left(\frac{1}{0.639 \text{ in.}^2/\text{tube}}\right)\left(\frac{144 \text{ in.}^2}{\text{ft.}^2}\right) = 430$ tubes

b. Number of tube passes and exchanger length

Information from Foster Wheeler Corporation's Bulletin OH-39-13 gave further evidence that the selection of 1-inch tubes was most practical. The very slight saving in first cost obtainable in selecting smaller tubes would hardly warrant their use when the trouble encountered in keeping a bundle with small tubes in satisfactory operating condition is considered.

Following the recommendations in this bulletin, tubes arranged on square pitch were selected for fouling service and easy cleaning.

Allowable distance between tubes = 0.25 inch

Using this spacing total cross-sectional area per tube = 1.56 in.^{2}

Total Heat Exchanger Cross-sectional Area = (1.56 in.^{2}/tube)(430 tubes/pass) = 670 $\cong$ 700 in.^{2}/pass

No. of Tube Passes = 8 (Arbitrarily selected to give reasonable length and diameter)

Heat Surface per Pass = $\frac{12{,}500 \text{ ft.}^2}{8}$ = 1560 ft.^{2}

Tube Length = (1560)/(430)(0.262)ft.^{2}/ft.) = 13.8 14 ft.

This tube length is in agreement with the value of 12 to 16 feet recommended in the bulletin mentioned above.

Total tubes per exchanger = (430)(8) = 3,440

c. Exchanger Diameter

Total cross-sectional area of exchanger = (700 in.^{2}/pass)8 = 5600 in.^{2}

Exchanger Diameter = $\sqrt{\frac{(5600)(4)}{\pi}}$ = 84.6 inches = 7 ft.

d. Number of Baffles and Baffle Spacing on Shell Side

Cross-flow type baffles are recommended and were selected in preference to orifice type baffles. The former require fewer baffles per exchanger and can be of heavier construction.

Minimum baffle thickness = 0.25 inch

No. of shell passes = 6

On this basis total free width of flow = 1.58 ft.

Water flow rate = $\left(\frac{4.11 \times 10^6}{4} \text{ lbs./hr.}\right)\left(\frac{1}{62.2 \text{ lbs./hr.}}\right)$

$\left(\frac{1}{360 \text{ secs./hr.}}\right)$ = 4.60 ft.^{3}/sec.

Allowable Water Velocity = 4 ft./sec.

Baffle Spacing = $(4.60\ ft.^3/sec.)\left(\frac{1}{1.58\ ft.\ width}\right)\left(\frac{1}{4}\ ft./sec.\right)$

= 0.727 ft. ≅ 9 inches

No. of Baffles per Shell Pass = $\frac{14\ ft.}{0.75}$ ≅ 18 Baffles

7. Pressure Drop Tube Side = 5 psi

8. Operating Pressure = 350 psia

C. Liquid Tank Separators

1. Material Balance

Input = w_5 = 5,308,669 lbs./hr.

Output = $w_{5b} + w_{5a} + w_{5d}$ = 99,205 + 69,464 + 5,140,000

= 5,308,669 lbs./hr.

2. Selection of Tank Size

Capacity of Separator = (5,308,669 lbs./hr.)/8.33 lbs./gal.

= 636,000 gals./hr.

In determining actual tank capacity, time must be allowed for filling, settling, and separating of liquids. In order to allow sufficient time for these operations, equal intervals of time were selected for each operation. Therefore, the total required capacity will be three times greater than that given above.

Total Capacity of Separator = (3)(636,000)

= 1,908,000 or 2,000,000 gals.

In order to determine the number and dimension of the tanks to be used, an economic balance was made by assuming various diameters and heights and calculating the respective costs. Metal wall thickness for a pressure of 300 psia was calculated as before.

Cost data were obtained from Vilbrandt pages 482-3.

The following tabulations indicate these calculations:

Diam. ft.	Gals/ft.[13]	Height	Capacity Gals.	No. Required	Volume of Metal per Tank Less Heads-ft^3	Wgt. Metal per Tank	Cost $/lb	Total Cost $
10	587.5	20	11,750	170	70.8	34,400	0.17	993,000
15	1321.9	30	39,700	51	239.6	116,000	0.12	710,000
20	2350.0	40	94,000	22	564.0	274,000	0.10	603,000
25	3672.0	50	183,600	11	1110.0	539,000	0.10	593,000

This economic balance indicated a minimum for 25-ft. diameter tanks 50-ft. high, for which a total number of 11 were required. Since there were three operations required, filling, emptying, and settling, it was decided to use three tanks with each active condenser or a total of 12 tanks.

3. Cost Calculations

Thickness of metal = 3.75 inches

Volume of metal in cylinder walls = 1110 ft.^{3}/tank
Volume of metal in semi-elliptical heads = 614 ft.^{3}/tank
Using mild carbon steel

$$\text{Total weight of metal} = (1110 + 614)\frac{\text{ft.}^3}{\text{tank}}\left(\frac{486\ \text{lb.}}{\text{ft.}^3}\right)$$

$$= 838{,}000\ \text{lbs./tank}$$

$$\text{Cost per tank} = (838{,}000\ \text{lbs/tank})(\$.10\ \text{per lb.})(1.08\ \text{Index Factor})^{(14)}$$

$$= \$90{,}000$$

These cost figures include the cost of roof, shell, manhole, nozzles, ladders, and freight for 100 miles and installation costs with the exception of foundations and painting.

Remaining Installation Cost = $9,000 (estimated)

Total Cost = ($99,000)(12) = $1,188,000

4. Operating Pressure - 300 psia

D. Liquid Recycle Pump

1. Pump and Motor Selection

Design was made for one pump per recycle cooler plus one extra spare or a total of 6. (4 in operation at any one time.)

$$\text{Liquid Rate} = (5{,}140{,}000 \text{ lbs./hr.})\left(\frac{1}{8.34 \text{ lbs./gal.}}\right)\left(\frac{1}{60 \text{ min./hr.}}\right)\left(\frac{1}{4 \text{ pumps}}\right)$$

$$= 2570 \text{ gpm}$$

Pressure Drop through Recycle Cooler = 5 psi

Pressure Drop Across Condenser Sprays = 25 psi

Friction Drop in Lines (estimate) = 30 psi

Total Pressure Drop = 60 psi

$$\text{Ideal Pump Work} = (60 \text{ lbs./in.}^2)(144 \text{ in.}^2/\text{ft.}^2)\left(\frac{1}{62.2 \text{ lbs/ft.}^3}\right)\left(\frac{1.283 \times 10^6 \text{ lbs.}}{\text{hr.}}\right)\left(\frac{1}{60 \text{ min./hr.}}\right)$$

$$= 2.98 \times 10^6 \text{ ft.-lbs./min.}$$

Ideal Horsepower = $2.98 \times 10^6/33{,}000 = 90.3$ Hp

Pump Shaft Efficiency = 80%

Motor Efficiency = 90%

Motor Horsepower = (90.3)/(0.8)(0.9) = 125 Hp

Pump Selected: Single stage, single suction, ball bearing, direct connected centrifugal pump.

Motor Selected: 125 Hp 220 volt AC 3-phase L.A. Type S (Vilbrandt Page 156 and 157)

2. Cost Calculations

Cost data from Vilbrandt page 194 Table 140

$$\text{Pump and Motor Cost} = (\$25/\text{Hp})(125 \text{ Hp})\underbrace{(1.32)}_{\text{Index Factor}}{}^{(14)} = \$4130 \text{ per pump}$$

Installation Cost = \$1770 per pump

Total Cost = (1770 + 4130)(6) = \$35,450

Power Required = (125 Hp/pump)(4 pumps)(0.746 KW/Hp)

= 373 KW

Power Cost = 373 KW x 0.0125 $/KW-hr. = $4.66 $/hr.

E. Pipe Diameters Required

All pipe diameters were selected from a nomograph included in reference.(16) For piping required for this section see flow diagram Figure 8 Page. All pipes for pressure of 300 psi.

1. Vapor Distributing Pipe for Condensers

Density = 0.588 lbs./ft.^{3} Rate = 55,200 ft.^{3}/min.

Actual Pipe Diameter = 80 inches

2. Inlet Pipes to Condensers

Density = 0.588 lbs./ft.^{3} Rate = 13800 ft.^{3}/min.

Actual Pipe Diameter = 35 inches

3. Vapor Line Condensers to Scrubber Line

Density $\cong$ 0.945 lbs./ft.^{3} Rate = 8150 ft.^{3}/min.

Actual Pipe Diameter = 35 inches

4. Liquid Line Condenser to Separators and Separator Distributing Line

Density $\cong$ 62.4 lbs./ft.^{3} Rate = 3402 gpm

Actual Pipe Diameter = 18 inches

5. Water and Water Soluble Chemicals Line from Each Separator

Density = 62.4 lbs./ft.^{3} Rate = 2,605 gpm

Actual Pipe Diameter = 15 inches

6. Water and Water Soluble Chemical Stream to Combine with Stream from Scrubber

Density $\cong$ 62.4 lbs./ft.^{3} Rate = 138.8 gpm

Actual Pipe Diameter = 4 inches

7. Recycle Line Pump to Intercooler, Separator to Pump, and Intercooler to Condenser

Density 62.4 lbs./ft.^{3} Rate = 2570 gpm

Actual Pipe Diameter = 15 inches

8. Cooler Water Lines (Inlet and Outlet to Sewer)

Density = 62.4 lbs./ft.^{3} Rate = 8230 gpm

Actual Pipe Diameter = 26 inches

9. Cooling Water Lines to and from Each Exchanger

Density = 62.4 lbs./ft.^{3} Rate = 2060 gpm

Actual Pipe Diameter = 14 inches

10. Oil and Oil Soluble Chemicals Lines from Each Separator

Density = 49.9 lbs./ft.^{3} Rate = 61.9 gpm

Pipe Diameter = 3 inch standard

11. Combined Oil Line to Refinery

Density = 49.9 Rate = 247.6 gpm

Actual Pipe Diameter = 5 inches

Numbers on lines in Figure 8 refer to the above numbers.

INSTRUMENTATION

I. Condensers, Separators, and Recycle Coolers

All cost data from Vilbrandt, pages 486 and 487

A. Condensers	Unit Cost $	Accessories $	Installation $	Total $
1. Five orifice or venturi meters and bodies, recording and controlling one on each vapor inlet line	600	240	240	5,400
2. Pneumatic Control system components (No compressor)	300	45	75	420
3. Five pressure recorders	150	45	60	1,275
4. Five temperature recorders and controllers.(6-point Electronic strip chart potentiometer) To record liquid outlet temperature and to control same by controlling flow of recycle coolant	675	270	338	6,415

B. Recycle Coolers				
1. Five orifice or venturi meters and bodies, recording only. On recycle inlet line.	500	200	200	4,500
2. Five temperature recorders and controllers. To record temperature of recycle stream leaving cooler and to control same by controlling water inlet flow to each cooler	675	270	338	6,415

C. Separators

The three separators with each condenser are to operate on an automatically timed and controlled cycle in which one is filling, one is settling and one is emptying.

Filling time = (183,600 gals./tank)$\left(\frac{1}{3402 \text{ gals/min.}}\right)$ = 54 mins.

Settling time = 54 mins.

Emptying time = 54 mins. First 1.3 minutes oil layer empties
Last 52.7 minutes water layer empties

	Unit Cost $	Accessories $	Installation $	Total $
1. 36 Pneumatic control systems	300	45	75	15,120
2. 36 Pneumatic transmission components	175	16	44	8,460
3. 4 Automatic timing devices	1000		500	4,500

D. Miscellaneous

1. One orifice flow recorder on combined water and water soluble chemicals line	500	200	200	900
2. One orifice flow recorder on combined oil and oil soluble chemicals line	500	200	200	900

Total Instrument and Accessories Cost = $41,660

Total Installation Cost = $12,645

Total Instrumentation Cost Including Installation = $54,305

MISCELLANEOUS COSTS

I. Condensers, Recycle Coolers and Separators

A. Piping Cost

25% of basic process equipment cost

Materials Cost = (0.25)($1,401,090) = $350,200

Installation Cost = (0.4)($350,200) = 140,200

B. Electric Wiring

1/2% of basic process equipment cost

Materials Cost = (0.005)($1,401,090) = 7,000

Installation Cost = (0.4)($7,000) = 2,800

C. Insulation, Fireproofing, and Paint

1% of basic process equipment cost

Materials Cost = (0.01)($1,401,090) = 14,000

Installation Cost = (0.4)($14,000) = 5,600

D. Depreciation Cost

10% per year

Total Equipment Cost = $2,196,205

Depreciation Cost = (0.1)($2,196,205) = $219,620 per year

E. Taxes, Rent, Insurance, Patent Rights and Maintenance = $219,620 per year

10% per year

F. Labor

1 pump man, 1 condenser man, 1 separator man

Total of three men per shift

Labor Cost = (3)(2.50 $/hr.)(8760 hrs./yr.) = $65,600 per year

Compound	Molecular Weight	Critical Temp. Tc °K	Critical Press. Pc Ats.	Vapor Heat Capacity at 300 psia Btu/lb.-mole °F. 600°F.	500°F.	400°F.	300°F.	200°F.	150°F.
CO_2	44	304	72.9	11.68	11.44	11.18	11.08	11.04	10.82
CO	28	134	34.6	7.37	7.28	7.24	7.23	7.21	7.20
H_2	2.0	33.2	12.8	7.01	7.00	6.98	6.95	6.92	6.90
N_2	28	12.6	33.5	7.18	7.15	7.14	7.13	7.13	7.13
H_2O	18	647	218	8.74	8.64	8.56	8.50	—	—
C_1-C_2	18	207	46.1						
C_3-C_4* (S)	51	297	39.5	33.05	29.14	29.1	27.1	25.3	24.5
C_3-C_4* (U)	49								
C_5-C_{10}* (S)	107	505	29.9	66.3	59.6	60.0	56.0	50.0	46.4
C_5-C_{10}* (U)	105								
C_{11}-C_{18}* (S)	205	703	16.5	136	127	118	105	94.4	86.1
C_{11}-C_{18}* (U)	203								
Alcohols	52	526	56.5	30.2	25.9	18.3			
Aldehydes	52	481	53.0	27.0	24.1	19.6	19.0	18.0	16.6
Ketones	60	235	47.0	32.8	30.3	27.2	24.1	20.1	18.6
Acids	64	322	57.2	31.5	26.1	24.8	21.2	17.3	14.5
Oil Soluble Chemicals	116	634	31.3	60.4	55.7	48.7	42.9	37.1	34.8

*S - saturated

*U - unsaturated

TABLE III

PHYSICAL PROPERTIES

Compound	Petroleum Characterization Factor	Normal Boil. Point °F.	Latent Heat of Vaporization at 1 ats. Btu/mole	Latent Heat of Vaporization 32°F. 300 psia	Vaporization Equilibrium Constants at 300 psia 150°F.	200	300	400	500	600
CO_2					∞	∞	∞	∞	∞	∞
CO					∞	∞	∞	∞	∞	∞
H_2					∞	∞	∞	∞	∞	∞
N_2					∞	∞	∞	∞	∞	∞
H_2O					0.0121	0.0499	0.34	.834	1.73	5.67
C_1-C_2		-244	3920		9.5	14	20	26	32	
C_3-C_4* (S) / C_3-C_4* (U)	12.6	-8	7900	5250	0.5	1.2	2.4	4.0	5.8	
C_5-C_{10}* (S) / C_5-C_{10}* (U)	12.35	228	16,500	5720	0.014	0.095	0.30	0.95	0.95	1.2
C_{11}-C_{18}* (S) / C_{11}-C_{18}* (U)	12.4	488	20,400	10,000	0.00286	0.00512	0.0165	0.0389	0.142	0.282
Alcohols	11.7	191	10,300	7,500	2.12	2.92	3.59	4.91	6.14	7.49
Aldehydes	10.3	120	9,700	7,500	2.08	2.80	3.38	4.74	5.98	7.29
Ketones	10.25	133	10,860	7,500	2.14	2.98	4.45	5.44	6.85	8.34
Acids	8.9	244	16,480	9,000	2.31	3.11	4.59	5.55	6.92	8.32
Oil Soluble Chemicals	9.7	347	17,300	10,000	.0408	.102	.322	.675	1.29	1.83

*S - saturated

*U - unsaturated

TABLE III - (Cont.)

PHYSICAL PROPERTIES

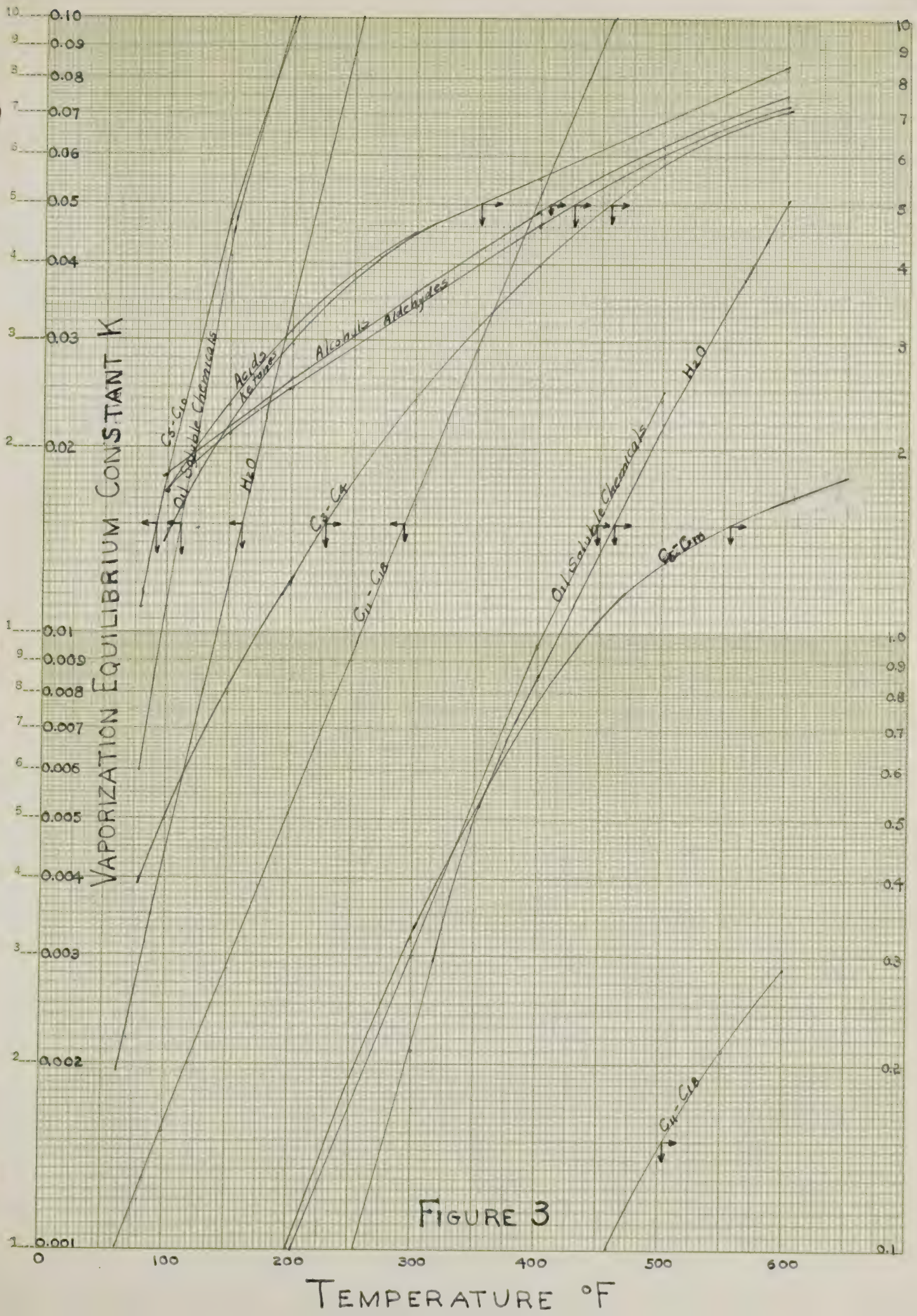

VAPORIZATION EQUILIBRIUM CONSTANT K
TEMPERATURE °F
0
100
200
300
400
500
600
0.10
0.09
0.08
0.07
0.06
0.05
0.04
0.03
0.02
0.01
0.009
0.008
0.007
0.006
0.005
0.004
0.003
0.002
0.001
1.0
0.9
0.8
0.7
0.6
0.5
0.4
0.3
0.2
0.1
C5-C10
Oil Soluble Chemicals
Acids
Ketones
Alcohols
Aldehydes
H2O
C3-C4
C11-C18
Oil Soluble Chemicals
H2O
C6-C10
C14-C18

FIGURE 3

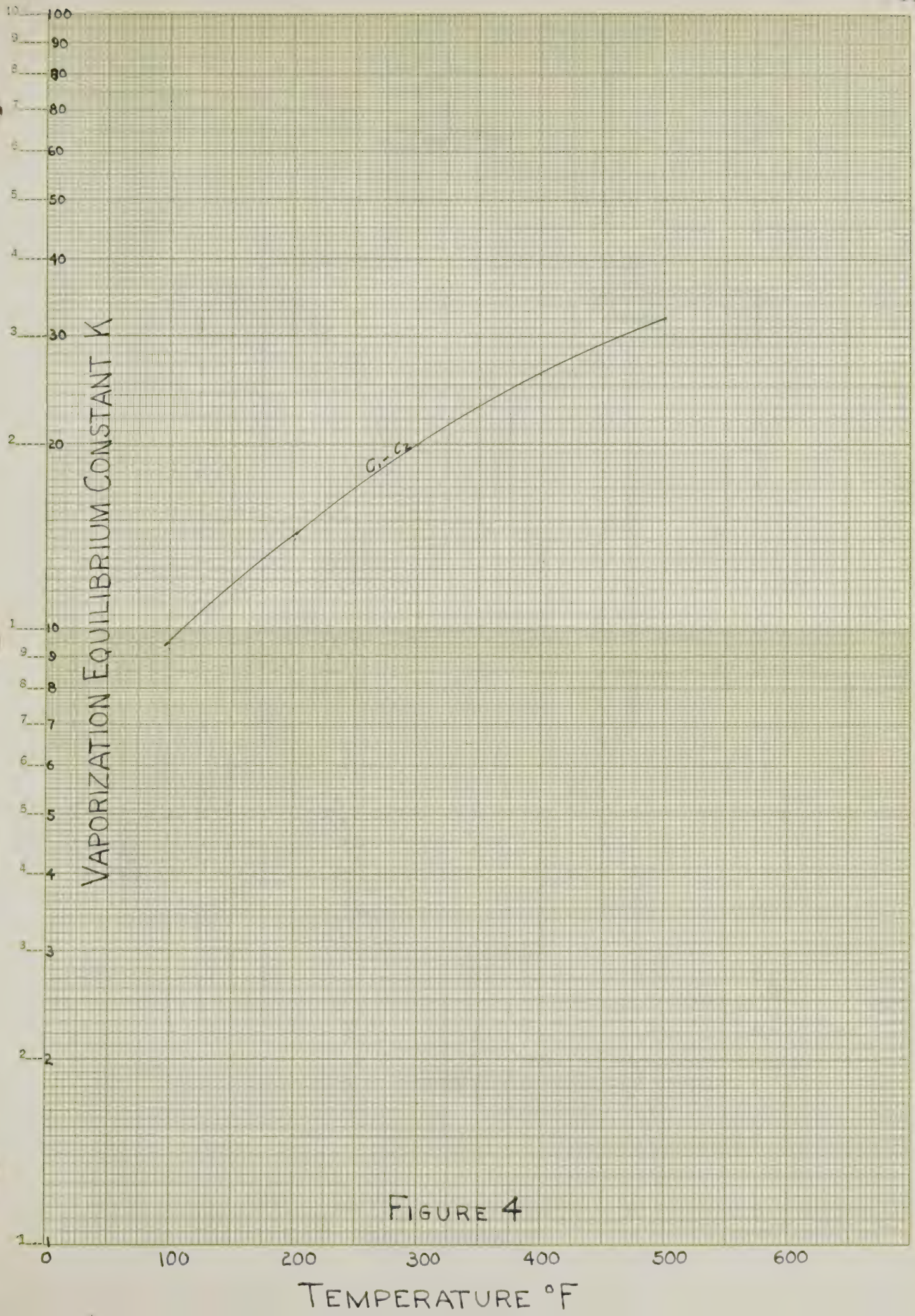
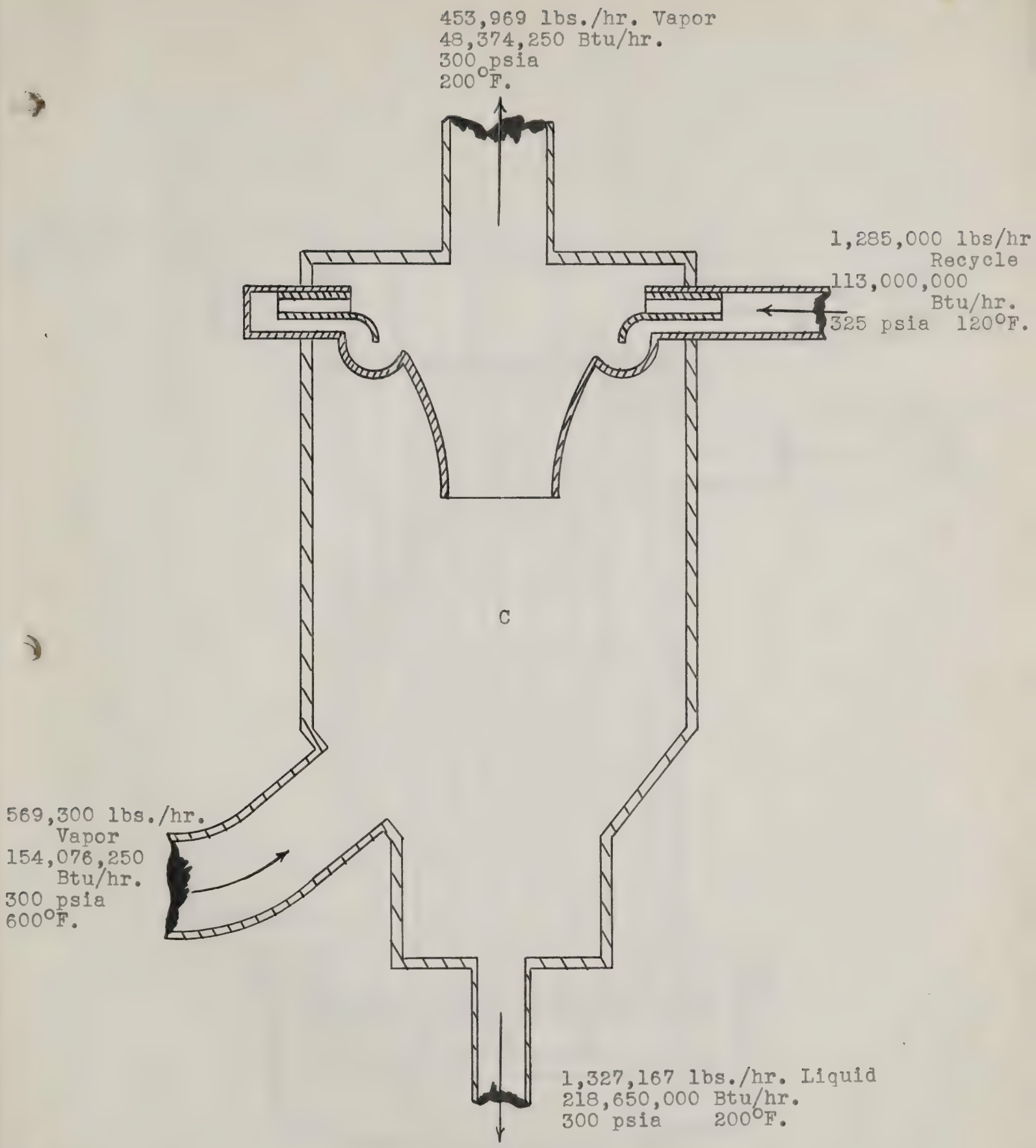


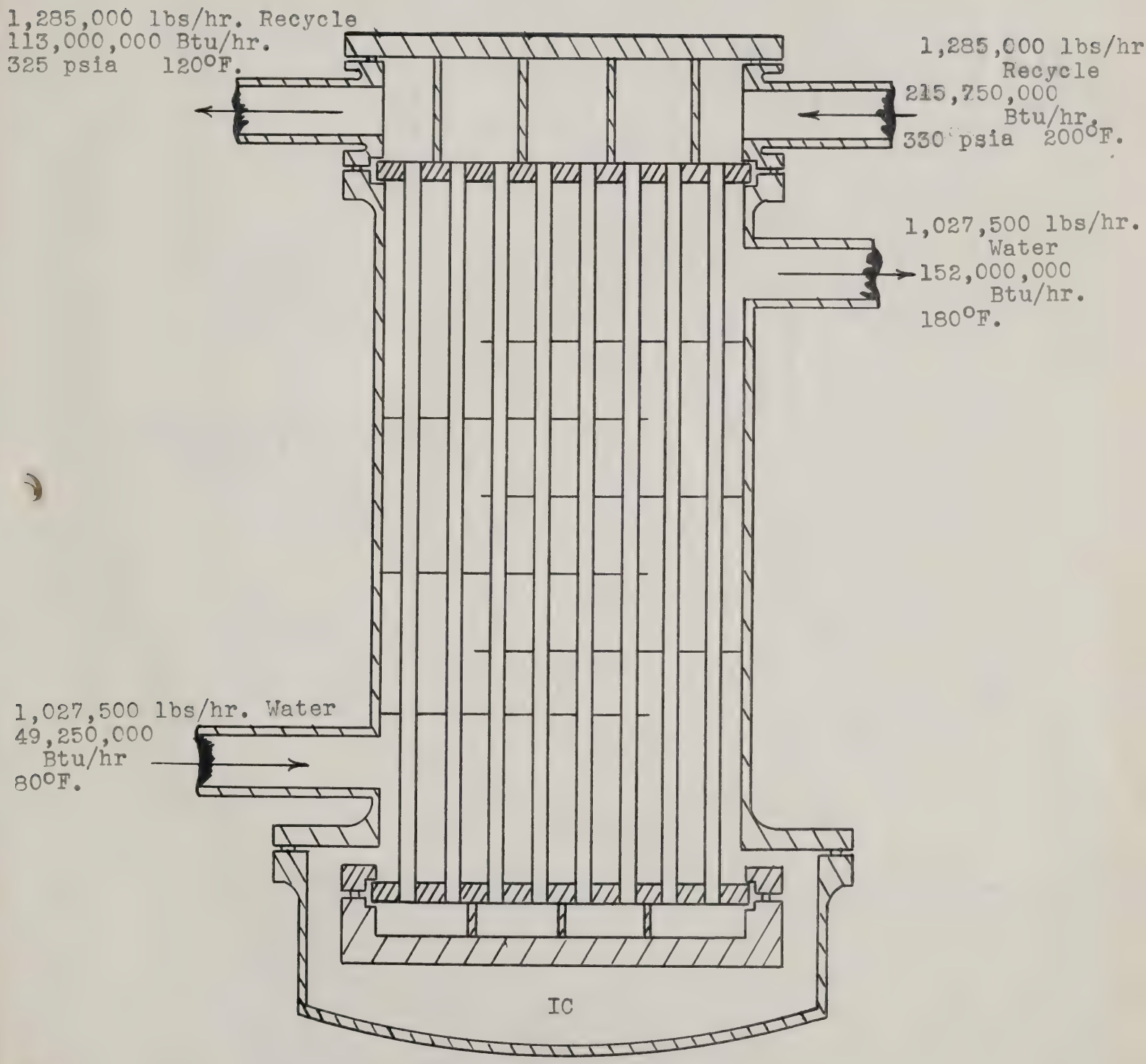
FIGURE 4

KEUFFEL & ESSER CO., No. 359-63
Semi Logarithmic, 2 cycles x 10 to the ½ inch
5th Lines accented
MADE IN U.S.A.



SPRAY CONDENSER

FIGURE 5



RECYCLE COOLER

FIGURE 6

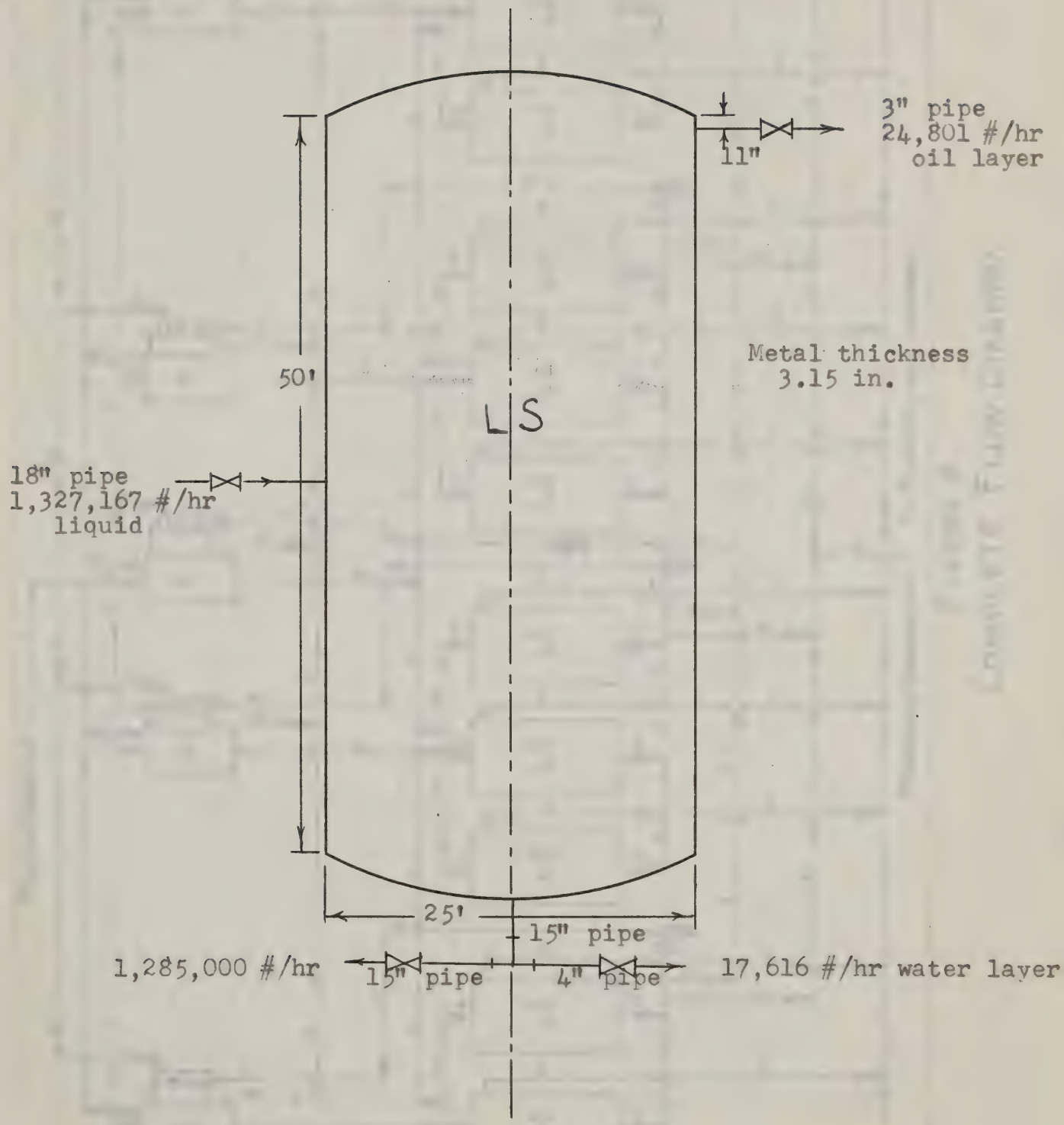
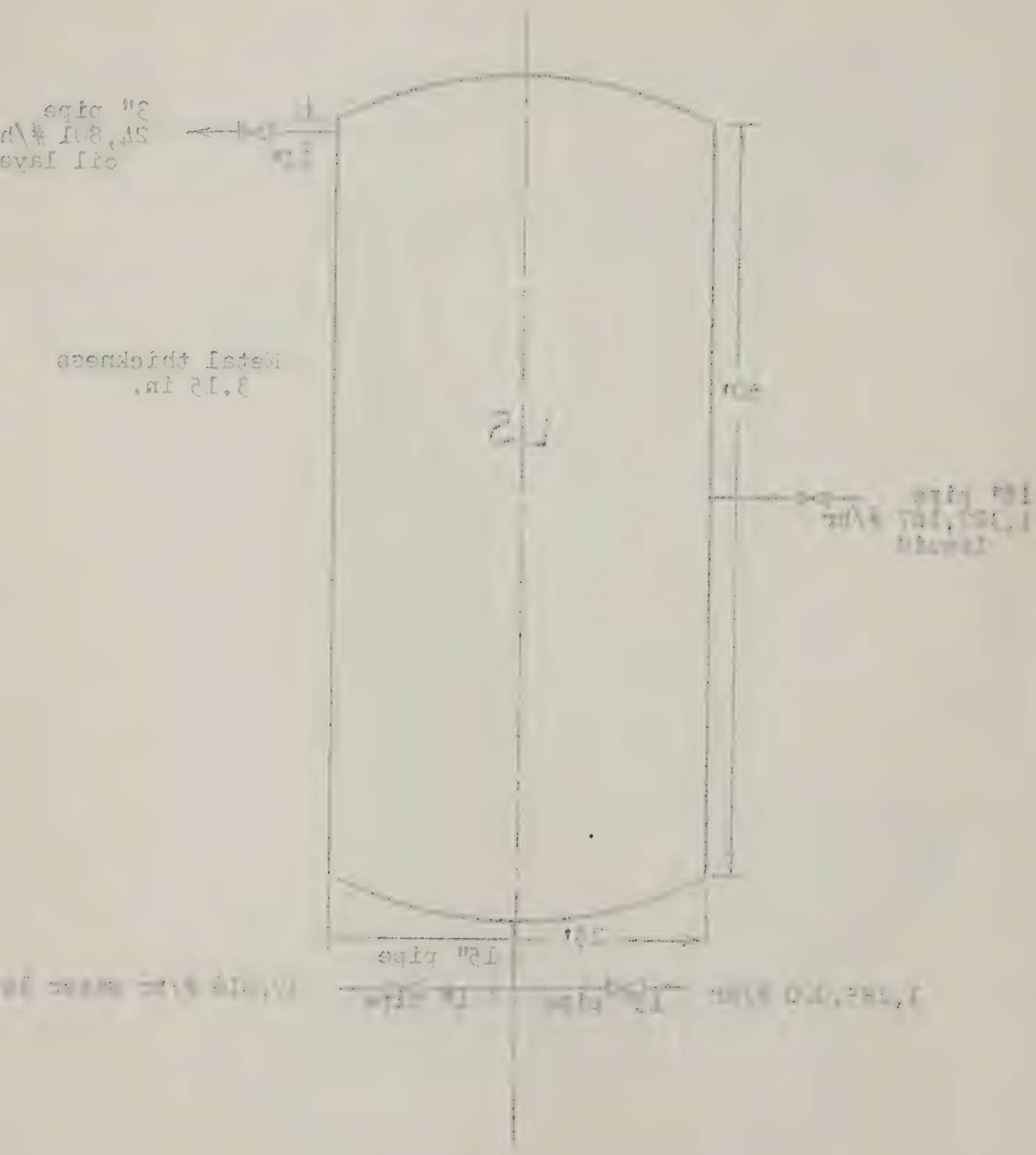


Fig. 7. Liquid Gravity Separators



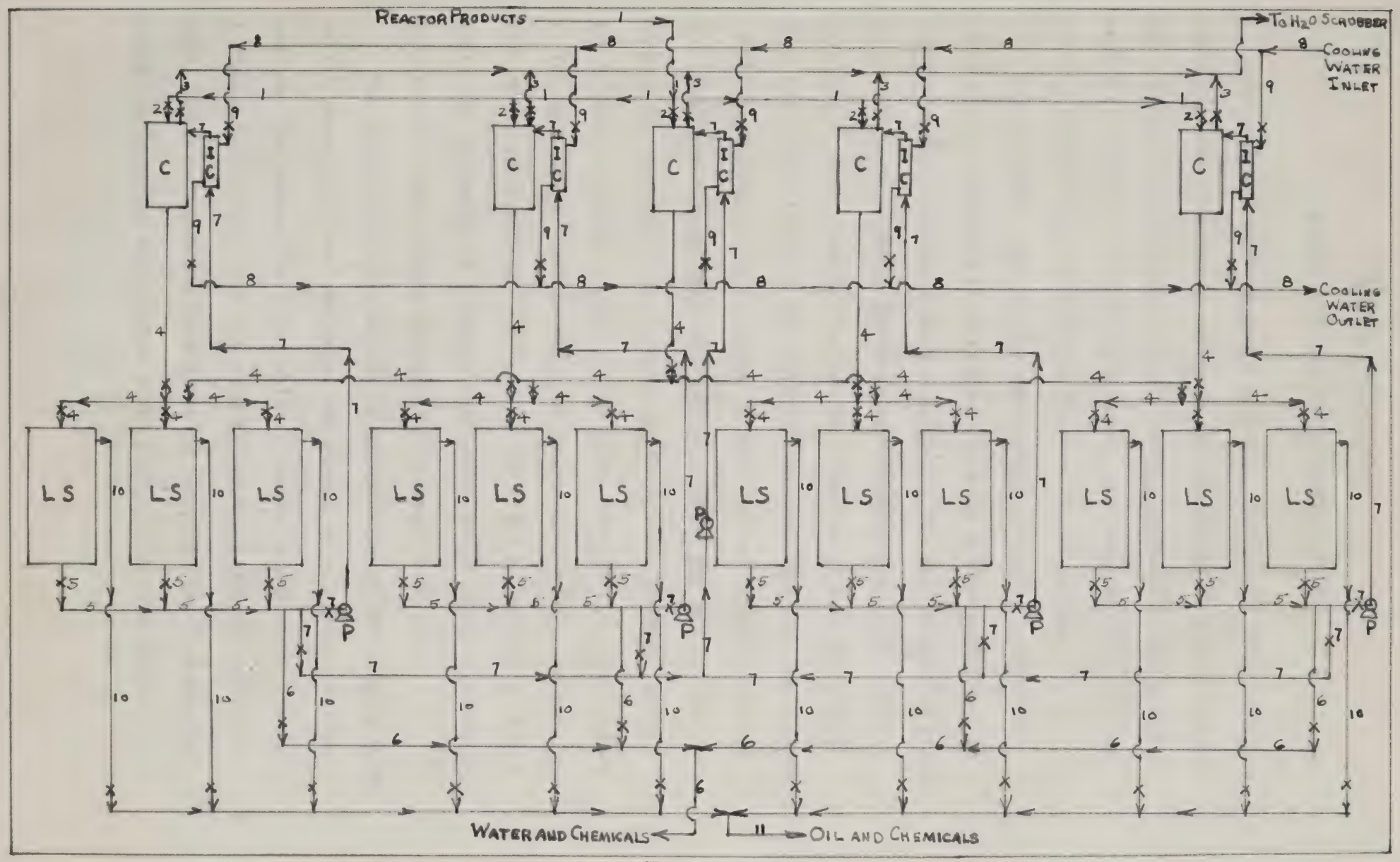


FIGURE 8

COMPLETE FLOW DIAGRAM

DESIGN CALCULATIONS
for
RECYCLE GAS COOLER AND DEHUMIDIFIER

The original intention was that this unit would be primarily a stripping tower to remove water soluble chemicals from the recycle gas stream. It soon became evident though, because of the almost negligible amount of chemicals present, that dehumidification and cooling would be the controlling factors in the design.

The only rate coefficient data available is for systems in which air is the inert carrier gas. For this system therefore, it was necessary to employ a unique approach. The method used is based upon an enthalpy driving force as described by Michley.[17] His development is applied to the dehumidification of air and hinges upon the assumption that the ratio of the heat transfer coefficient to the mass transfer coefficient is equal to the specific heat of the humid gas stream. This assumption is valid for air but not necessarily valid for other inert carrier gases. However, since the average molecular weight of the gases involved in this problem is fairly close to that of air, (22 as compared to 29), it was felt that the error involved in accepting the above assumption as valid would not be nearly so great as that which would result from using rate coefficients defined specifically for use with air.

The problem then became one of calculating an equilibrium curve of the enthalpy of the saturated gas stream as a function of temperature. The method for doing this is adequatly described below. A straight operating line was assumed and the required number of equilibrium contacts was determined by 'stepping off' between the operating line and equilibrium curve over the desired range .

In lieu of tray efficiency data based upon enthalpy contacts, a tray efficiency of 20 per cent was assumed. It is felt that this is sufficiently low enough to compensate for any erroneous steps which may have been taken in the determination of the required number of contacts. The calculations involved are given in the following pages.

INCOMING GAS STREAM: $T = 200^{o}F$; $P = 300$ psia; $H = 193{,}497{,}000$ BTU/hr; (relative to 32 ^{o}F)

Comp.	Mol/hr	#/hr	C_p(32-200) BTU/Mol-^{o}F	Latent Ht. BTU/Mol	C_pX(mol/hr) (for dry gas)
CO_2	18520	815000	11.00		203720
CO	16000	448000	7.23		115700
H_2	32620	65300	6.90		225200
N_2	6990	195800	7.20		50300
H_2O*	3150	56700	8.60	Stm.Tbl.	
C_{1-2}	2986	53700	11.30		33750
C_{3-4}	1232	61600	24.50		30200
C_{5-10}	824	87300	46.00		37900
C_{11-18}	14	2855	86.00		1203
Alc.*	239	12450	20.00	7500	
Ald.*	48	2490	16.60	7500	
Ket.*	43	2580	18.60	7500	
Acid*	101	6460	14.50	9000	
Oil Sol.	49	5640	34.00		1665
	82816	1815875			699688

*Classed as wet gases

82816 total mols/hr
3582 wet mols/hr
79234 dry mols/hr

Avg. C_p(for dry gas) $= \frac{699688}{79234} = 8.95$ BTU/mol^{o}F

1,734,195 #/hr of dry gas entering

Avg. Mol. Wt. of dry gas $= \frac{1734195}{79234} = 21.9$

Enthalpy of entering stream $= \frac{193497000}{1734195} = 111.6$ BTU/# of dry gas

The desired equilibrium curve is a plot of the enthalpy of the saturated gas stream in BTU/# of dry gas versus temperature. The amount of water vapor present at any temperaure is determined from data out of the steam tables. To determine the concentration of the

water soluble chemicals at any temperature, it was assumed that the chemicals are dissolved linearly with temperature through the tower; i.e. their concentration at 200° is that given above while at 85° (temperature of the exit gas stream) no chemicals are present in the gas stream, and at a mid-point temperature they are present at one half of their original concentration. Thus the amount of each chemical present at any temperature is easily determined from the linear graphs in Fig. 9.

The enthalpy of the saturated gas stream is then calculated as the sum of the sensible heat of the dry gases plus the latent heats of the chemicals at 32° and their sensible heats from 32° to the desired temperature plus the enthalpy of the water vapor present as obtained from the steam tables. The enthalpy of the dry gas stream at 200° was determined by subtracting from the enthalpy of the total stream that of the wet components as follows.

enth. of H_2O vapor:	56700 #/hr X 1146 BTU/# =	65,000,000
enth. of Alc.vapor:	7500 BTU/mol X 239 Mol =	1,791,000
	20 BTU/mol°F X 168°F X 239 Mol =	803,500
enth. of Ket.vapor:	7500 X 43 =	322,300
	18.6 X 168 X 43 =	134,300
enth. of Ald.vapor:	7500 X 48 =	360,000
	16.6 X 168 X 48 =	136,700
enth.of Acid vapor:	9000 X 101 =	910,000
	14.5 X 168 X 101 =	246,000
		69,771,300

193,497,000
69,771,300
123,725,700 BTU/hr, Enthalpy of dry gas @ 200°F

The pounds of water vapor present are determined from steam table data in the following manner: p_{H_2O} = Vapor Pressure of Water from Table.

$$\frac{p_{H_2O}}{P_{Total}} = \frac{mols_{H_2O}}{mols_{gas} + mols_{H_2O}} \qquad m_g \cdot \frac{p_{H_2O}}{P_T} = m_{H_2O}\left(1 - \frac{p_{H_2O}}{P_T}\right)$$

$$18 \times m_{H_2O} = \frac{18 \cdot m_g \cdot p_{H_2O}}{P - p_{H_2O}} = \text{pounds of water vapor}$$

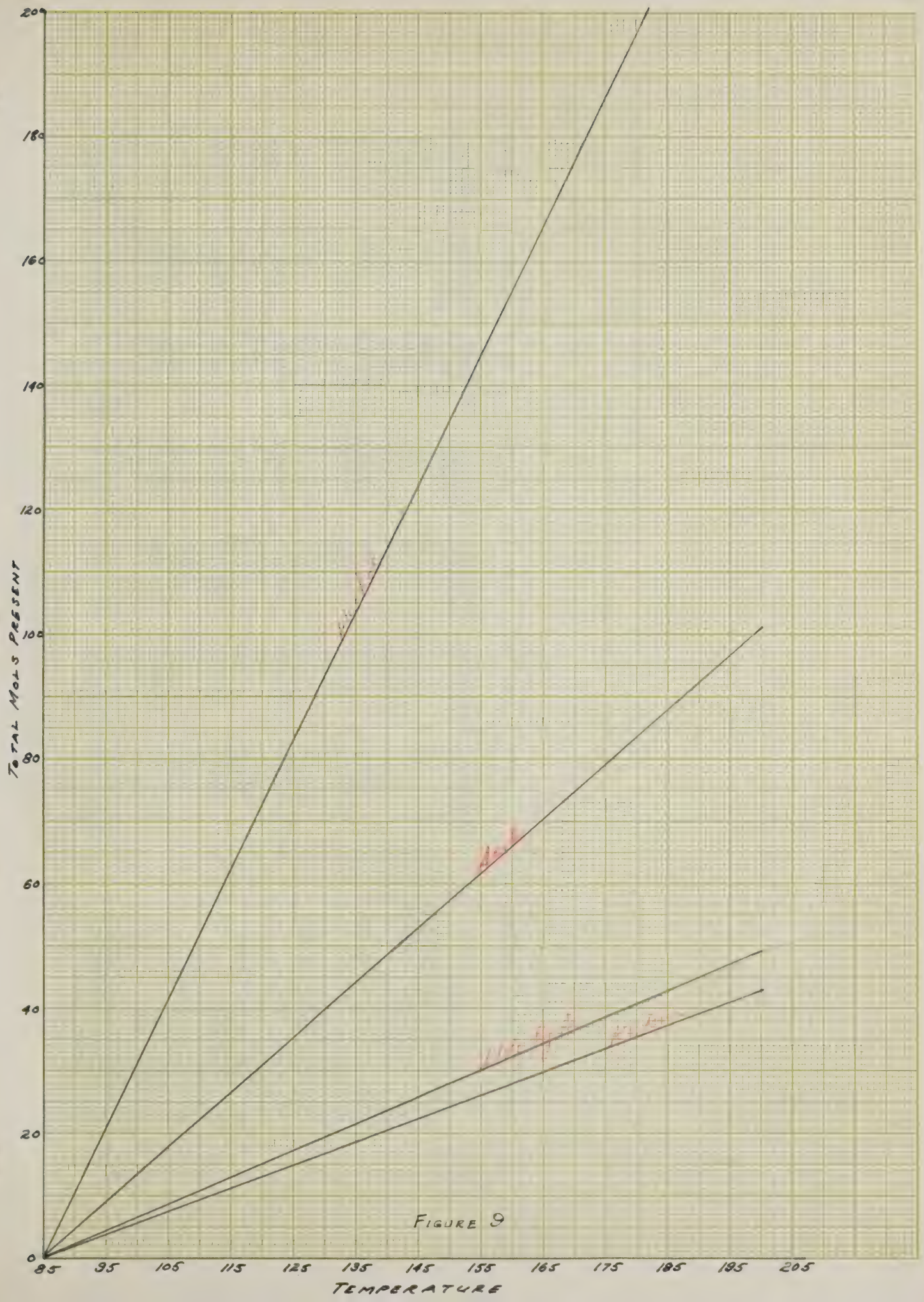


FIGURE 9

KEUFFEL & ESSER CO.
NO. 358-11. 10 × 10 to the half inch, 5th lines accented
Engraving, 7 × 10 in.
MADE IN U.S.A.

TABULATED CALCULATIONS

(1)	Temperature	180	160	140	120	100	90	80
(2)	79235·8.95(200-T)	14000000	28000000	42000000	56000000	70000000	77000000	84000000
3)	123,725,700-(2)	109725700	95725700	81725700	67725700	53725700	46725700	39725700
4)	mols alc.	196.5	155	114	72.5	31	10	0
5)	7500·(4)	1473000	1162000	855000	543500	232500	75000	0
6)	(4)·20·(T-32)	581000	396500	246200	127200	42150	116000	0
7)	mols ald.	40	32	23.5	15	6.5	2.4	0
8)	7500·(7)	300000	240000	176100	112500	48700	18000	0
9)	(7)·16.6·(T-32)	97250	68000	42150	21870	7340	2310	0
10)	mols ket.	35.5	28	20.5	13	5.5	2.0	0
11)	7500·(10)	266000	210000	153700	97500	41250	15000	0
12)	(10)·18.6·(T-32)	97600	65000	41100	21200	6950	2160	0
13)	mols acid	83	66	48	31	13	4.3	0
14)	9000·(13)	747000	594000	431500	279000	117000	38700	0
15)	(13)·14.5·(T-32)	178000	122200	75200	39500	12820	3620	0
16)	$p(H_2O)$	7.51	4.74	2.89	1.69	0.95	0.70	0.507
17)	79234·(16)·18/300-(16)	36150	22600	13700	7980	4460	3285	2380
18)	H_{vap} @ T (Stm.Tbl.)	1138	1130	1122	1114	1105	1101	1097
19)	(17)·(18)	41100000	25550000	15390000	8900000	4925000	3620000	2610000
20)	Σ	154965550	124133400	99136650	77867990	59159400	50512090	42335700
21)	i_g^* (20)/1734195	89.4	71.55	57.1	44.85	34.1	29.1	24.4

The equilibrium curve (i_g^* versus T) is shown plotted in Fig.10. Cooling water is available at 80°F and it was arbitrarilly assumed that the out-let gases would be cooled to 85° with the water being heated to 190° to give Δt's of 5° and 10° at the top and bottom of the tower respectively. As was pointed out above, a straight operating line was assumed. It was determined that 7.6 theoretical trays would be required for the desired dehumidification by simply stepping off the equilibrium contacts between the operating and equilibrium lines.

The necessary amount of cooling water was determined from a heat balance in the following manner:

Heat taken up by water = Heat given up by gases

$$= 111.6 - 26.75 = 84.85 \text{ BTU/\#dry gas}$$

$$\text{or, } 84.85 \times 1{,}734{,}195 = 14.7 \times 10^7 \text{ BTU/hr}$$

Water goes from 80° to 190°, $\Delta t = 110°$

therefore, $\dfrac{14.7 \times 10^7}{110} = 1.339 \times 10^6$ # water per hr.

$$\text{or, } \frac{1.339 \times 10^6}{8.33} = 160.8 \times 10^3 \text{ gal/hr}$$

$p_{H_2O} = 0.596$ @ 85°F

therefore,

$$\text{Water in out gases} = \frac{79234 \cdot 0.596 \cdot 18}{300 - 0.596} = 2800 \text{ \#/hr}$$

Water in entering gases = 56700 #/hr

Water removed 56700 - 2800 =	53900	#/hr
Alcohol removed	12450	
Aldehyde removed	2490	
Ketone removed	2580	
Acid removed	6460	
Total removed from gas	77880	#/hr
Cooling water in	1339000	
Total liquid stream out	1416880	#/hr

FIGURE 10

KEUFFEL & ESSER CO.
NO. 358-11. 10 × 10 to the half inch, 5th lines accented
Engraving, 7 × 10 in
MADE IN U.S.A.

CONCENTRATION OF CHEMICALS IN LIQUID STREAM:

$$\frac{12{,}450 \times 100}{1{,}416{,}880} = 0.880 \text{ per cent alcohol by weight}$$

$$\frac{2{,}490 \times 100}{1{,}416{,}880} = 0.176 \text{ " " aldehyde " "}$$

$$\frac{2{,}580 \times 100}{1{,}416{,}880} = 0.182 \text{ " " ketone " "}$$

$$\frac{6{,}460 \times 100}{1{,}416{,}880} = 0.456 \text{ " " acid " "}$$

1.694 per cent total chemicals

Justification for the assumption that all of the water soluble chemicals will be removed from the gas:

Raoult's Law is valid for the solute in dilute solution. Therefore,

$$p = P_o x \qquad y = \frac{p}{P} = \frac{P_o x}{P}$$

where, y = mol fraction of solute in vapor
x = mol fraction of solute in liquid
p = partial pressure of solute in vapor
P_o = vapor pressure of pure solute
P = total gas pressure

In order to include a maximum factor of safety, the calculations are based upon the supposition that all of the chemicals are dissolved in the inlet water stream; this gives values of x which are higher than would actually be encountered in the tower.

Component	P_o (@ 80°F)	Total mols	x	y
ROH (ethyl)	15 psia	239	0.00423	0.000212
RCHO (acet.)	30	48	0.000845	0.0000845
RCOR (acetone)	30	43	0.00754	0.0000754
RCOOH (acetic)	0.26	10	0.00178	0.00000154

It is apparent that the mol fraction of the chemicals in the vapor is negligible.

TOWER DIMENSIONS:

The allowable vapor velocity was determined from the equation:

$$V_{allowable} = 0.225\sqrt{\frac{\rho_L - \rho_G}{\rho_G}}$$

At bottom of tower: ($t_g = 200°$, $t_l = 190°$)

$$\rho_G = \frac{21.9}{359} \times \frac{492}{660} \times \frac{300}{14.7} = 0.941 \ \#/ft^3$$

$$\rho_L = 60.35 \ \#/ft^3$$

At top of tower: ($t_g = 85°$, $t_l = 80°$)

$$\rho_G = \frac{21.9}{359} \times \frac{492}{545} \times \frac{300}{14.7} = 1.14 \ \#/ft^3$$

$$\rho_L = 62.2 \ \#/ft^3$$

Top of tower:

$$V = 0.225\sqrt{\frac{62.2 - 1.14}{1.14}} = 1.642 \text{ ft/sec}$$

$$\frac{79234}{3600} \times 359 \times \frac{545}{460} \times \frac{14.7}{300} = 423.00 \text{ ft}^3\text{/sec of dry gas}$$

$$\frac{2800}{18 \times 3600} \times 359 \times \frac{545}{460} \times \frac{14.7}{300} = \frac{0.84}{423.84} \text{ ft}^3\text{/sec water vapor total}$$

$$\frac{423.9}{1.642} = 258 \text{ ft}^2 \text{ area required at top}$$

Bottom of tower:

$$V = 0.225\sqrt{\frac{60.35 - 0.941}{0.941}} = 1.782 \text{ ft/sec}$$

$$\frac{82816}{3600} \times 359 \times \frac{660}{460} \times \frac{14.7}{300} = 535 \text{ ft}^3\text{/sec total gas}$$

$$\frac{535}{1.782} = 300 \text{ ft}^2 \text{ area required at bottom}$$

Therefore, use 300 ft^2

$$D = \sqrt{\frac{4 \times 300}{\pi}} = 19.5 \text{ ft}$$

Use three towers in parallel to cut down the necessary diameter

$$A = 100 \text{ ft}^2 \qquad D = \frac{19.5}{1.732} = 11.25 \text{ ft}$$

For operation at 300 psia, ratio of wall thickness to diameter for mild steel is 0.025.

therefore,

$$\text{Shell thickness} = 5.625 \times 0.025 \times 12 = 1.69''$$

Tray Spacing: (Chem. Eng. Prog., 45, 719, (1949))

$$\text{'F' factor} = u_{vap}\sqrt{\rho_G} = 1.782\sqrt{0.941} = 1.73$$

This 'F' factor is beyond the range of the graphical correlation given in the above article. It appears reasonable however, to assume a tray spacing of 2 feet.

To compensate for errors which may have been made in the determination of the theoretical contacts required, estimate a modest tray efficiency of 20 per cent.

therefore,

$$\frac{7.6}{0.2} = 38 \text{ actual trays required per tower}$$

$$\text{tower height} = 38 \times 2 = 76 \text{ feet}$$

Tower Requirements:

Number needed	3
Height	76'
Inside diameter	11.25'
Shell thickness	1.69"
Plates required per tower	38
Material	Mild Steel

ECONOMIC PIPE SIZES:

Economic pipe diameters were determined from a nomograph by Colburn[6] which relates the proper diameter with flow rate, density and viscosity.

(Letters used refer to the lines as designated in Fig. 11)

Line	Flow Rate	Density, #/ft^3	Economic Diameter
A	2680 gpm	62.2	15 inches
B = D = E	894 "	"	9.3
C	1796 "	"	13
J	2840 "	60.3	15.5
H	1893 "	"	13.5
F = G = I	947 "	"	9.5
a	32600 cfm	0.931	67
b = d = e	10867 "	"	40
c	21732 "	"	55
j	25450 "	1.138	60
f = g = i	8483 "	"	36
h	16967 "	"	50

PRESSURE DROP THRU TOWER:

Estimate drop of 6 inches of water per plate.

$$\text{Total drop} = 38 \times \frac{6}{12} \times \frac{62.4}{144} = 8.22 \text{ psi}$$

Insofar as the calculations for this unit were concerned, this pressure drop was considered as being negligible. For the overall energy balance of the separation unit however, it is taken into consideration.

GAS STREAM LEAVING TOWER: 85°F, H = 46,400,000 BTU/hr

Component	Mol/hr	#/hr
CO	18,520	815,000
CO	16,000	448,000
H	32,620	65,300
N	6,990	195,800
H O	155.6	2,800
C	2,986	53,700
C	1,232	61,600
C	824	87,300
C	14	2,855
Oil Sol.	49	5,640
	78,391	1,736,995

COOLER-DEHUMIDIFIER FLOW SHEET

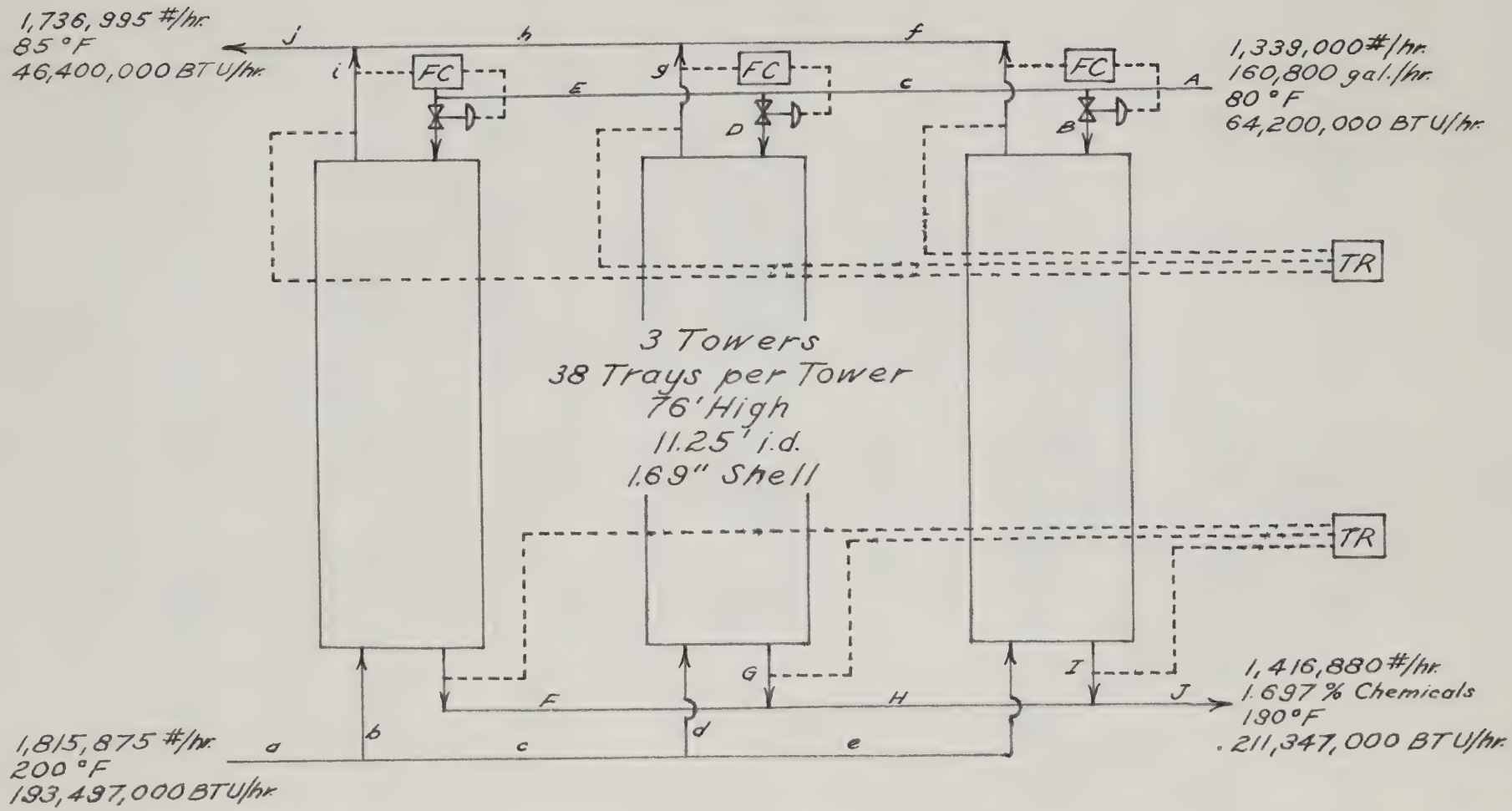


Fig. 11

K.E.K. 6/50

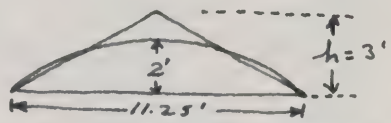
COST ESTIMATION:

Towers,
For the shell, base cost on 0.096 \$/# for mild steel, having a density of 486 #/ft^3

Shell volume $= \pi DtH = \pi\left(11.25 + \frac{1.69}{12}\right)\frac{1.69}{12} \times 76 = 383 \text{ ft}^3$

Volume of flat base $= \pi R^2 t = \pi\left(\frac{11.25}{2} \quad \frac{1.69}{12}\right)^2 \times \frac{1.69}{12} = 14.75 \text{ ft}^3$

Volume of dome top



$$V \doteq \frac{1}{3}\pi h(2rt + t^2)^2 \doteq \frac{\pi}{3} \times 3 \times \left[\frac{2 \times 11.25 \times 1.69}{12} + \left(\frac{1.69}{12}\right)^2\right]^2$$

$$\doteq 33.8 \text{ ft}^3$$

Total volume of steel per shell = 431.55 ft^3

Cost = 431.55 X 486 X 0.096 = \$20,150

Tray Costs,
\$10.00 to \$15.00 per ft^2 (Vilbrandt, P.473) (valid for 1947)
Cost per ft^2 would be lower for larger plates, but, considering the increases in costs between 1947 and the present time, it was decided that \$15.00 per ft^2 would be the more correct value to use.

Tray area $= \pi \times (5.625)^2 = 95 \text{ ft}^2$
Total tray cost per tower = 95 X 38 X 15 = \$54,000

Total cost per tower, shell and trays = \$74,150

Assume 40 per cent for fabrication, including tray supports, nozzles, flanges, man-holes, etc.

Fabricated tower cost = 74,150 X 1.4 = \$103,900

Assume installation cost is $\frac{1}{0.7}$ X fabricated cost (Contest Prob.)

Installed cost per tower = $\frac{103,900}{0.7}$ = \$148,000

Total installed cost for three towers = \$444,000

Piping Costs,
Assume piping equipment cost to be 25 per cent of basic equipment cost. Use the above factor for installation cost.

Equipment for the three towers = 103,00 X 3 X .25 = \$77,925
Installed piping cost = $\frac{1}{0.7}$ X 77,925 = \$111,321

Instrument Costs, (Vilbrandt, P.487)

Require: 2, 3-point temperature recorders
3, Flow rate recorders and controllers
(temperature activated control)

Item	Flow Inst.		Temp. Rec.	
Base instrument	3 X 160	$480	2 X 800	$1600
Accesories	0.95 X 480	455		
Installation	0.50 X 480	240	0.5 X 1600	800
Thermocouples	3 X 120	360		
Pneumatic control system	3 X 300	900		
Accesories	0.15 X 900	135		
Installation	0.45 X 900	405		
		$2975		$2400

Total instrument cost = $5375.00

Electric Wiring and Lighting,

½ per cent of basic equipment cost

or, 0.005 X 3 X 103,900 = $1,559

installation 0.4 X 1559 = 624

Total $2,183

Paint, Fire-proofing, Insulation, etc.

½ per cent of basic equipment cost

or, 0.005 X 3 X 103,900 = $1,559

Total Installed Costs for Entire Unit,

3 plate towers	$444,000
Piping	111,321
Instrumentation	5,375
Electric wiring and lighting	2,183
Paint, etc.	1,559
	$564,438

Cost per Hour for Dehumidification Unit, (exclusive of insurance)

Based upon amortization period of 10 years

Amortization $\frac{564,438}{10 \times 24 \times 365}$ = $6.44 per hr

Water @ $0.04 per 1000 gal = 160.8 X 0.04 = 6.43

Labor, One man @ $2.50 per hr 2.50

TOTAL HOURLY COST $15.37

BIBLIOGRAPHY

1. Edmister, W. C., "Applications of Thermodynamics to Hydrocarbon Processing", Reprinted from Pet. Ref. Nov., 1948 to Dec., 1949, Pg. 64, Figure 69

2. Sherwood, T. K., "Absorption and Extraction", 100, (Figs. 28 and 29), 1st Ed. McGraw-Hill Book Co., Inc., New York City, New York, (1937)

3. Hougen, O. A. and Watson, K. M., "Chemical Process Principles", Part II, 666, Fig. 152, John Wiley and Sons, Inc., New York City, New York, (1948)

4. Hougen, O. A. and Watson, K. M., "Chemical Process Principles" Charts", Fig. 152, John Wiley and Sons, Inc., New York City, New York, (1948)

5. Hodgman, C. D., "H ndbook of Chemistry and Physics", Chemical Rubber Publishing Co., Cleveland, Ohio, 25th Ed., (1941)

6. Perry, J. H., "Chemical Engineer's Handbook", McGraw-Hill Book Co., New York City, New York, 2nd Ed., (1941)

7. Ibid 4 Figs. 5a, 63, 64, 65

8. Ibid 1 Pgs. 5 and 6, Figs. 37 and 38

9. Ibid 4 Figs. 67, 108

10. Ibid 4 Figure 68

11. McAdams,"Heat Transmission", McGraw-Hill Book Co., New York City, New York

12. Vilbrandt, "Chemical Engineering Plant Design", McGraw-Hill Book Co., New York City, New York, 3rd Ed., (1949)

13. Perry, J. H., "Chemical Engineer's Handbook", 123, McGraw-Hill Book Co., New York City, New York, 2nd Ed., (1941)

14. Chem. Eng. 57, 123, (1950)

15. Perry, J. H., "Chemical Engineer's Handbook", 1988, McGraw-Hill Book Co., New York City, New York, 2nd Ed. (1941)

16. Ind. Eng. Chem. 32, 1249, 1940

17. Mickley, Chem. Eng. Prog., Dec. 1949

Oil Absorber and Stripper

Summary

Absorber – 2 units – Diameter 11.28' Cost $146,[illegible]
Height 70.0'

Stripper 2 units Diameter 29.6' Cost 190,[illegible]
Height 37

Auxiliaries Cost 143,[illegible]

Total $479,[illegible]

Utilities

Steam Required 180,000 lb/hr.

Water Required

Power Required [illegible] lb/hr.

Operation

Economic Considerations

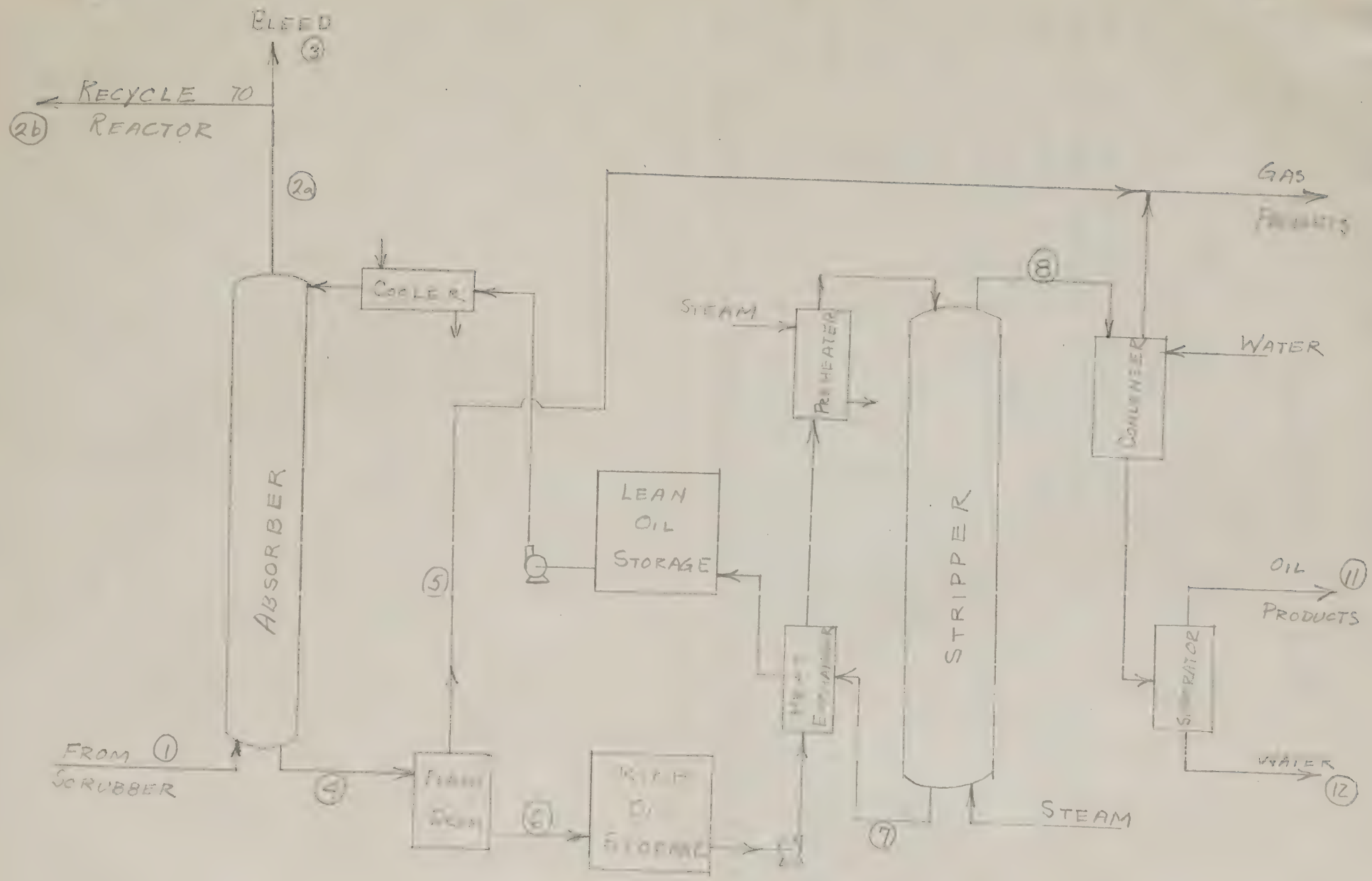
The oil absorber serves a two-fold purpose. The first is to remove the hydrocarbon products which were not condensed. The second purpose is to control the amount of methane in the recycle stream. It is impractical to allow the methane to build up in the recycle stream until the methane produced is all removed in the bleed since the amount produced is almost ten times as great as the amount of nitrogen introduced. Absorption of the heavier hydrocarbons is relatively easy while methane is absorbed with difficulty. The absorption of the methane is then the controlling factor. The economics of this situation favor the removal of as little methane as possible since the methane is of little value. The maximum amount of methane which may be returned to the reactor was estimated at 2360 mole/hr. requiring an absorption efficiency for this component of 50%.

Absorber oil is used in this apparatus. Due to the difficulty in stripping out some of the heavier hydrocarbons, they are allowed to accumulate with oil until they are stripped out at the same rate they are generated in the reactor. This causes some loss of product to the recycle

This loss is comparatively small.

The operation of the absorber at high pressure is favored for two reasons. First, it reduces the cost of recompressing the large volume of recycle gas, and second, the absorption is favored by pressure. The stripping process, however, is favored by low pressure. The cost of raising the pressure of the liquid oil is comparatively small. Further, a considerable quantity of light hydrocarbons is removed from the absorber oil when it is flashed to a lower pressure.

In the stripper, the removal of methane from the absorber oil is again the controlling factor since it cannot be allowed to accumulate in the absorber oil to any great extent. Since stripping is favored by high temperatures, the feed is preheated to 150°.



FLOW DIAGRAM OF ABSORBER AND STRIPPER

Summary of STREAM Compositions

Basis: 1 hour - Composition in mols

Comp.	1.	2a	2b	3	4	5	6	7	8
CO_2	18 520	18 520	17746	774	—	—	—	—	—
CO	16 000	16 000	15312	688	—	—	—	—	—
H_2	32 620	32 620	31 259	1361	—	—	—	—	—
N_2	6 990	6990	6 699	291	—	—	—	—	—
H_2O	151	151	144	7	—	—	—	—	10 000
C_1-C_2	5 347	2462	2360	102	2900	435	2465	10	2455
C_3-C_4	1232	—	—	—	3009	—	3009	1777	1232
C_5-C_{10}	824	—	—	—	71 000	—	71 000	70176	824
$C_{11}-C_{15}$	14	—	—	—	3060	—	3060	3046	14
Oil Sol. C.	48	—	—	—	1660	—	1660	1612	48
Abs. Oil	—	—	—	—	396 379	—	396 379	396 379	—
Total	81,796	76 743	73 520	3223	478 008	435	477 573	473.000	14 573

CALCULATIONS

Absorber

Allowable vapor velocity

$$u = K_v \sqrt{\frac{\rho_1 - \rho_2}{\rho_2}} \qquad \text{(Perry, p 1449)}$$

Density of Absorber = 1.85 F = 7.04 lb/gal (Kirkbride, IEC 35, 1242 (1943))

$$\rho_1 = \frac{7.04}{8.33} \times 62.4 = 52.7 \text{ lb/cu ft.}$$

Average Molecular Weight of Gas = 19.7

$$\rho_2 = \frac{19.7}{359} \cdot \frac{315}{14.7} \cdot \frac{492}{545} = 0.106 \text{ lb/cu ft.}$$

For 2 inch seal and 18" spacing; $K_v = 0.12$

Correction factor (0.75) applied as recommended by Perry.

$$u = (0.75)(0.12)\sqrt{\frac{52.7 - 0.106}{0.106}} = 2.0 \text{ ft/sec}$$

Area Required

$$\frac{80{,}000}{3600} \cdot \frac{359}{14.7} \cdot \frac{545}{315} \cdot \frac{492}{2.0} = 199 \text{ sq ft.}$$

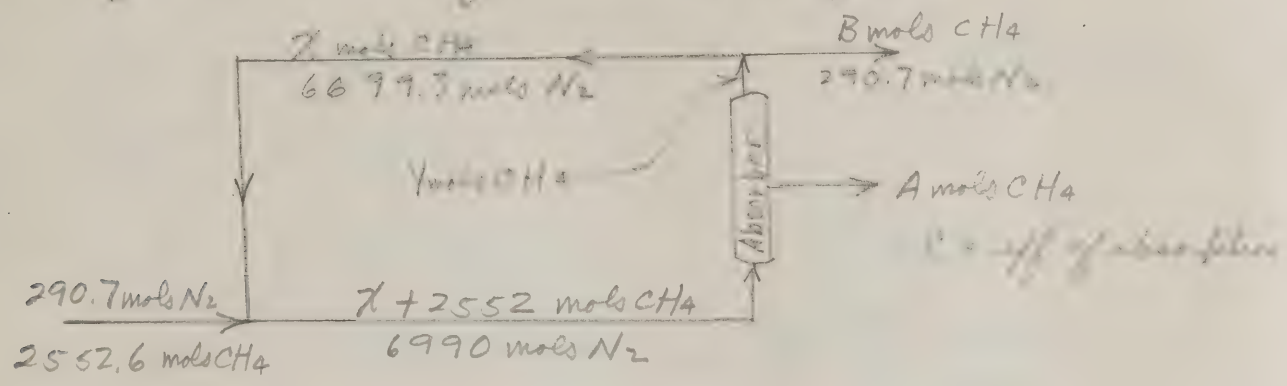
For n identical units:

$$D = \sqrt{\frac{4A}{\pi n}}$$

n	D	cost/tray for 1 unit	cost/tray for n units
1	15.92	$ 2800	$ 2800
2	11.25	1700	3400
3	9.2	1200	3600
4	7.97	1000	4000

Two units of 11.28' diameter were chosen due to the pressure of the gases.

Calculation of the amount of CH_4 in recycle:



$$Y = (1-\varepsilon)(X + 2552.6) = \frac{6990}{6699.3} X$$

or $$\varepsilon = \frac{2552.6 - 0.041 X}{2552.6 + X}$$

For 50% recovery, $X = 2360$ mols/hr.

$X + 2552.6 = 4912.6$ mols/hr.

Mols removed by absorption $= (0.50)(4912.6) = 2456.3$ mols/hr.

Mols in bleed $= B = (2360)\left(\frac{290.7}{6699.3}\right) = 102$ mols/hr

---

For dilute mixtures such as this, the Kremser equation (Sherwood, p 114) may be applied:

$$Eff = \frac{Y_1' - Y_0'}{Y_1' - m X_0'} = \frac{\left(\frac{L}{mG}\right)^{n+1} - \left(\frac{L}{mG}\right)}{\left(\frac{L}{mG}\right)^{n+1} - 1} = \frac{A^{n+1} - A}{A^{n+1} - 1}$$

where $A = \frac{L}{mG}$

For CH_4 $m = 11.5$ at 300 psi = 85°F

Kirkbride 35
1242 (1943)

let $\dfrac{A^{n+1} - A}{A^{n+1} - 1} = 0.5$

$A^{n+1} - A = 0.5A^{n+1} - 0.5$

$G = 81,746$ mol/hr

$A^{n+1} = 2A - 1$

Then:

n	$A = \frac{L}{mG}$	L	Initial cost
4	0.519	487,500	$ 90,[illegible]
5	0.509	478,000	113,500
6	0.504	473,000	136,200
7	0.502	472,000	158,900
8	0.501	471,000	181,600

Overall plate efficiency was assumed to be 15% (Sherwood, p 121)

Cost per actual tray = $3400

Cost per theoretical tray = $\dfrac{3400}{0.15}$ = $22,700

Number of actual trays = $\dfrac{6}{0.15}$ = 40

Tower height = $\dfrac{6}{0.15} \times 1.5 = 60 + 10 = 70$ ft.

Diameter = 11.5 ft.

A tower having 6 theoretical trays was selected since it has the lowest cost for a comparatively low oil rate. The oil rate must be kept as low as possible to make the stripping as easy as possible

Stripper

Estimated Equilibrium constants - (Sherwood p 106)
at 1 atmosphere

$T \rightarrow$	85°	150°	200°
$C_1 - C_2$	195	275	335
$C_3 - C_4$	6.8	14	19
$C_5 - C_{10}$	0.075	0.40	0.65
$C_{11} - C_{18}$	0.018	0.16	0.42
O.S.C.	0.35	1.00	1.50

Equilibrium flash Calculation - absorber oil is flashed to atmospheric pressure. A considerable amount of $C_1 - C_2$ is vaporized. Amounts of other components in vapor are low since they have relatively low values of K

Thus: $v_i = \dfrac{E_i}{1 + \frac{L}{K_i V}} = V$ for equilibrium flash

$$V = \frac{2900}{1 + \frac{478{,}000}{(195)(V)}}$$

$V = 435$ mols/hr.

Second Equilibrium flash - The rich oil is then heated to 150°F before being fed to stripper. More $C_1 - C_2$ is vaporized before the liquid comes in contact with the stripping steam. Thus another flash calculation must be made.

	mols	K_{150}	for $L=800$	$\frac{f}{1+L/KV}$ $L=820$	$L=815$
C_1-C_2	2955	275	775	787	785
C_3-C_4	1232	14	28	29	29
C_5-C_{10}	824	0.40	–		
$C_{11}-C_{18}$	14	0.16			
OSC	48	1.00			
			803	816	814

Thus effective liquid feed to stripper is:

C_1-C_2	1640
C_3-C_4	1203
C_5-C_{10}	824
$C_{11}-C_{18}$	14
OSC	48
Abs. Oil	473000
Total =	476,729

In the stripper, the gases are not been ruptures. Furthermore the temperature is not constant throughout the length of the tower.

The Kremser equation is applied to stripping as

$$\text{eff} = \frac{X_{n+1} - X_1}{X_{n+1}} = \frac{(S_e)^{n+1} - (S_e)}{(S_e)^{n+1} - 1}$$

where $S = \frac{KG}{L}$

$$S_e = \sqrt{S_n(S_1+1) + 0.25} - 0.5$$

Edminster – I.E.C. 35, 837 (1943)

Calculation of Se for $C_1 - C_2$ for various steam rates, Estimate exit temp of gas = 200°F.

Mols of steam	G_1	K_1	S_1	G_n	K_n	S_n	S_e
1200	1200	335	0.347	4929	275	0.285	0.281
3000	3000		2.115	6729		0.389	0.710
5000	5000		3.52	8729		0.505	1.390
10,000	10,000		7.05	13,729		0.795	2.08

Calculation of Se for $C_{11} - C_{15}$, the heaviest component.

Mols of steam	G_1	K_1	S_1	G_n	K_n	S_n	S_e
3000	3000	0.42	0.00265	6729	0.16	0.00227	0.00227
5000	5000		0.00442	8729		0.00294	0.00289
10,000	10,000		0.00885	13,729		0.00463	0.00458
20,000	20,000		0.01770	23,729		0.00798	0.00799

Heavier components can not be stripped out without allowing some build up in the absorber and thus causes a loss in the absorber. Since the efficiency of stripping of the heavier components is 1.0, $Y_0' = m X_1'$ in absorber. In the stripper, where Se is small, Efficiency = Se. Thus the Percent loss in the absorber may be estimated:

In absorber: $$Y_0 = \frac{Y_0'}{m} = \frac{(14)(\%\text{loss})}{(81,746)(0.00255)} \quad \text{for } C_{11} - C_{15}$$

or in stripper: $$X_1 = X_0' = 0.000672(\%\text{loss})$$

$$(X_{n+1} - X_1) = \frac{14}{473,000} = 0.0000296$$

$$Eff = \frac{X_{n+1} - X_1}{X_{n+1}} = Sc$$ when Sc is small

% loss	(X_1)	X_{n+1}	Eff	Sc
1	0.000672	0.0007013	0.0418	0.0418
5	0.00336	0.00339	0.00839	0.00829
10	0.00672	0.00674	0.00394	0.00394

a loss of 10% (1.4 mols/hr) may be tolerated without any appreciable build up of $C_{11} - C_{18}$ in recycle.
This would require about 10,000 mols of steam per hour
at this steam rate, it may be estimated from Sherwood p85
that for Sc = 2.08 for CCl_2, 6 theoretical plates would be required. Thus for $C_1 - C_2$

$$Eff = \frac{(2.08)^7 - (2.08)}{(2.08)^7 - 1} = 99.5\%$$

Dimensions

Max G = 13,729 mols/hr. at 220°F

Max U = 2 ft/sec,

$$Area = \frac{13,729}{3600} \Big| \frac{359}{} \Big| \frac{680}{492} \Big| \frac{}{2} = 948 \text{ ft}^2$$

n	D	cost/plate 1 unit	cost/plate for n units
1	34.8 ft	$15,000	$15000
2	24.6	5,500	11,000
3	20.1	4,400	13,200
4	17.4	3,500	14000

minimum cost is for two units 24.6' in diameter

actual trays = $\frac{6}{0.33}$ = 18

Height = (18)(1.5) + 10 = 37 ft'

Cost = (18)(11,000) = $198,000

Efficiencies for all components (In Stripper)

Comp.	G_1	K_1	S_1	G_n	K_n	S_n	S_e	Eff.
$C_1 - C_2$	10,000	335	7.05	13,729	275	0.795	2.08	99.5
$C_3 - C_4$		19	0.400		14	0.405	0.403	40.2
$C_5 - C_{10}$		0.65	0.01369		0.40	0.01155	0.0116	1.16
$C_{11} - C_{18}$		0.42	0.00885		0.16	0.00463	0.00458	0.458
O.S.C.		1.50	0.0316		1.00	0.0289	0.0289	2.89

Efficiencies for all components in absorber

α = 473,000

Comp	G	K	A	Eff.
C_1	81,746	11.5	0.504	50%
C_2		1.8	3.22	100
$C_3 - C_4$		0.42	13.80	100
$C_5 - C_{10}$		0.0097	598.0	100
$C_{11} - C_{18}$		0.00255	2270	100
OSC		0.034	170.5	100

Estimation of buildup – oil stream

Comp	Mols stripped	Eff	Mols entering	Mols leaving
$C_1 - C_2$	1640	99.5	1650	10
$C_3 - C_4$	1203	40.2	2920	1777
$C_5 - C_{10}$	824	1.16	71,000	70,176
$C_{11} - C_{18}$	14	0.458	3060	3046
OSC	48	2.89	1660	1612

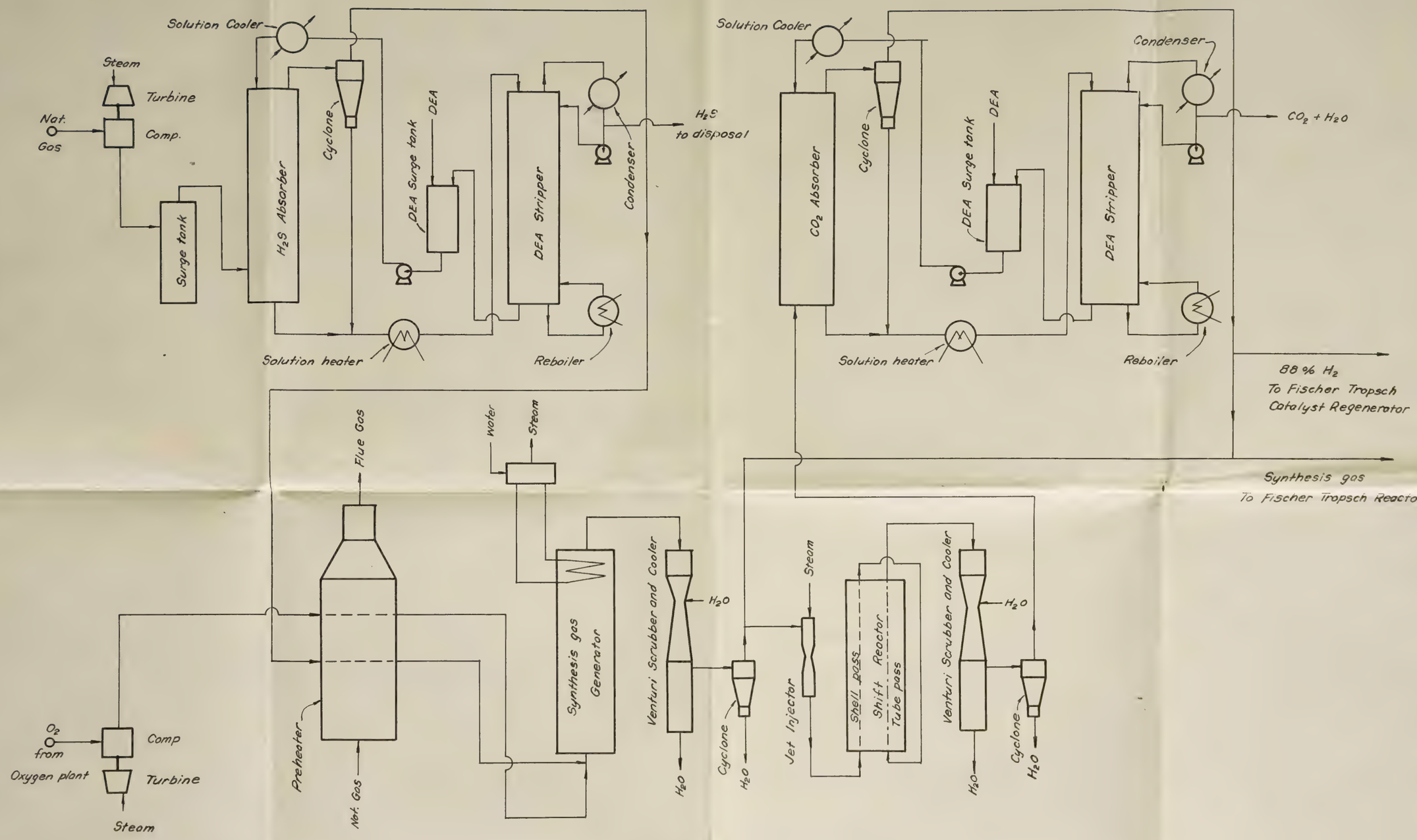
Composition of the oil must be corrected according to the [illegible] computation. This causes a slight build up of products in the recycle stream, whereas $y_0' = m x_0'$ for all except C

$$y_0' = \frac{mG_0}{L_0} b_0'$$

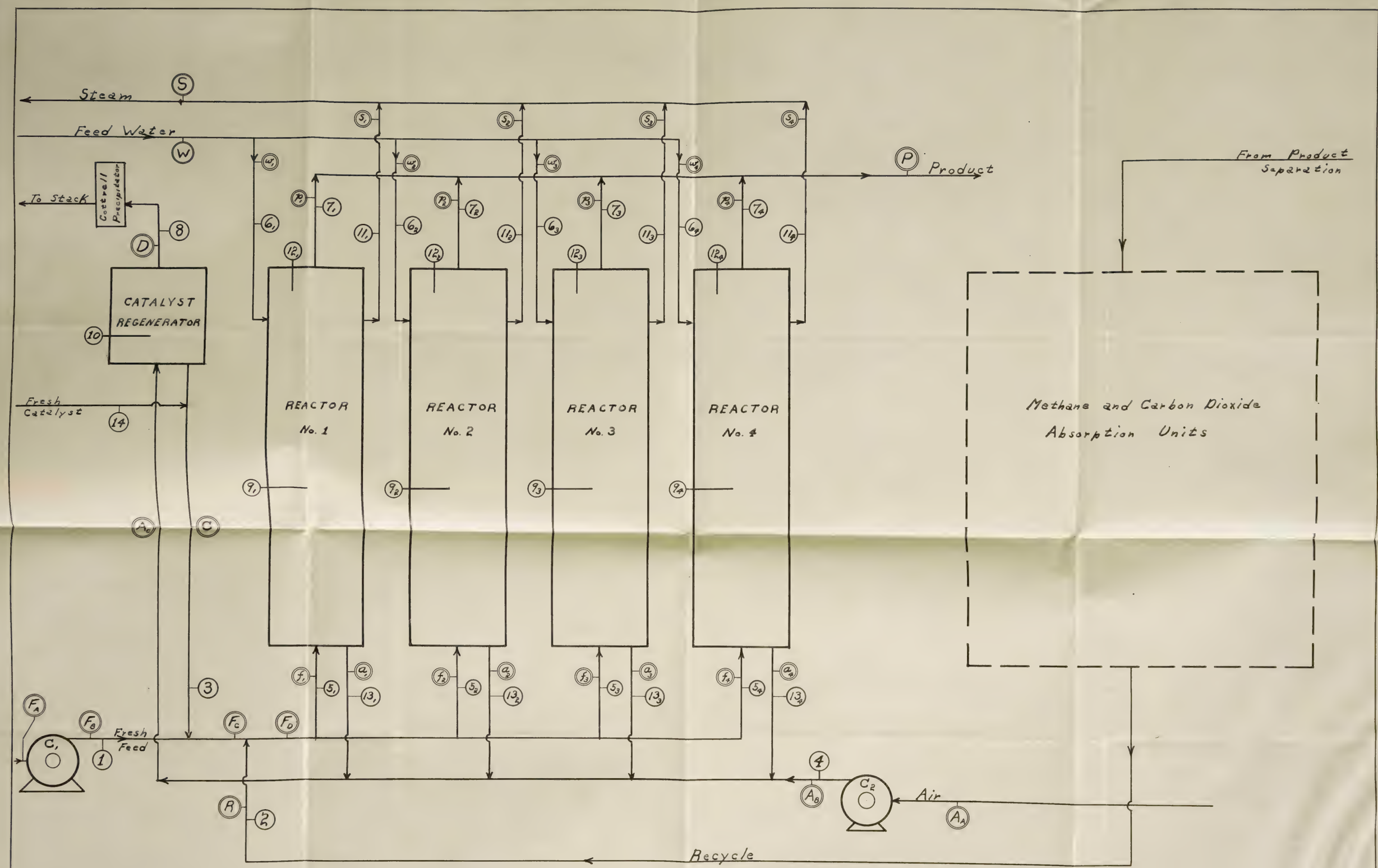
	b_0'	m	loss in oil
C_1	10		2462
C_2	–	1.5	—
$C_3 - C_4$	1777	0.42	127.2
$C_5 - C_{10}$	70,176	0.0097	117.2
$C_{11} - C_{18}$	3046	0.00255	1.34
OSC	1612	0.034	7.45

It is seen that the build up of all the components except methane is negligible.

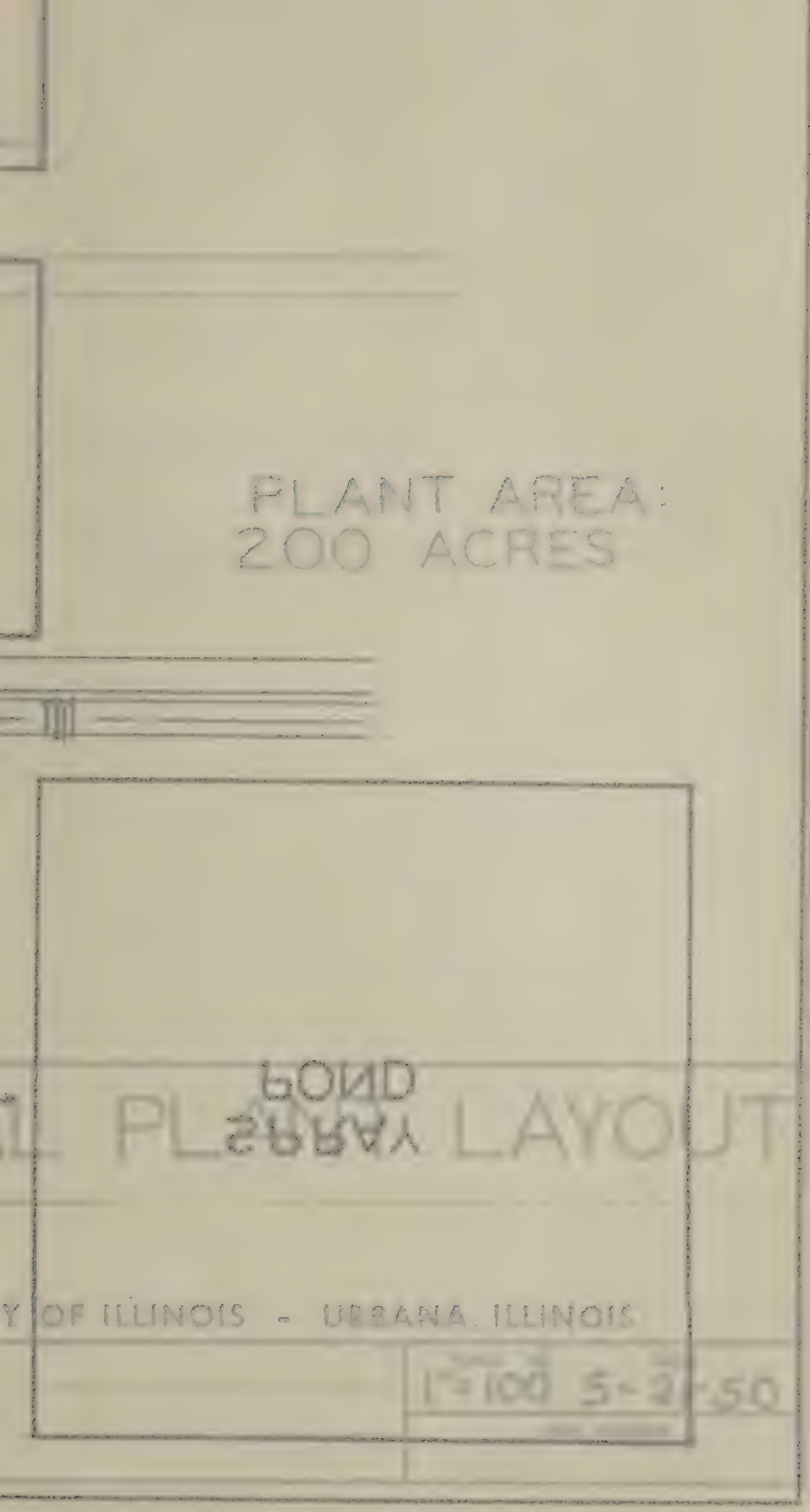
SYNTHESIS GAS PLANT



FLOW SHEET FOR SYNTHESIS GAS PLANT



DRAWN BY	USED ON	FLOWSHEET for Reactor and Catalyst Regeneration System for Proposed Fischer-Tropsch Plant
CHECKED BY		[redacted]
		UNIVERSITY OF ILLINOIS - URBANA, ILLINOIS
SUBMITTED		SCALE / DATE
EXAMINED		
APPROVED		PART NUMBER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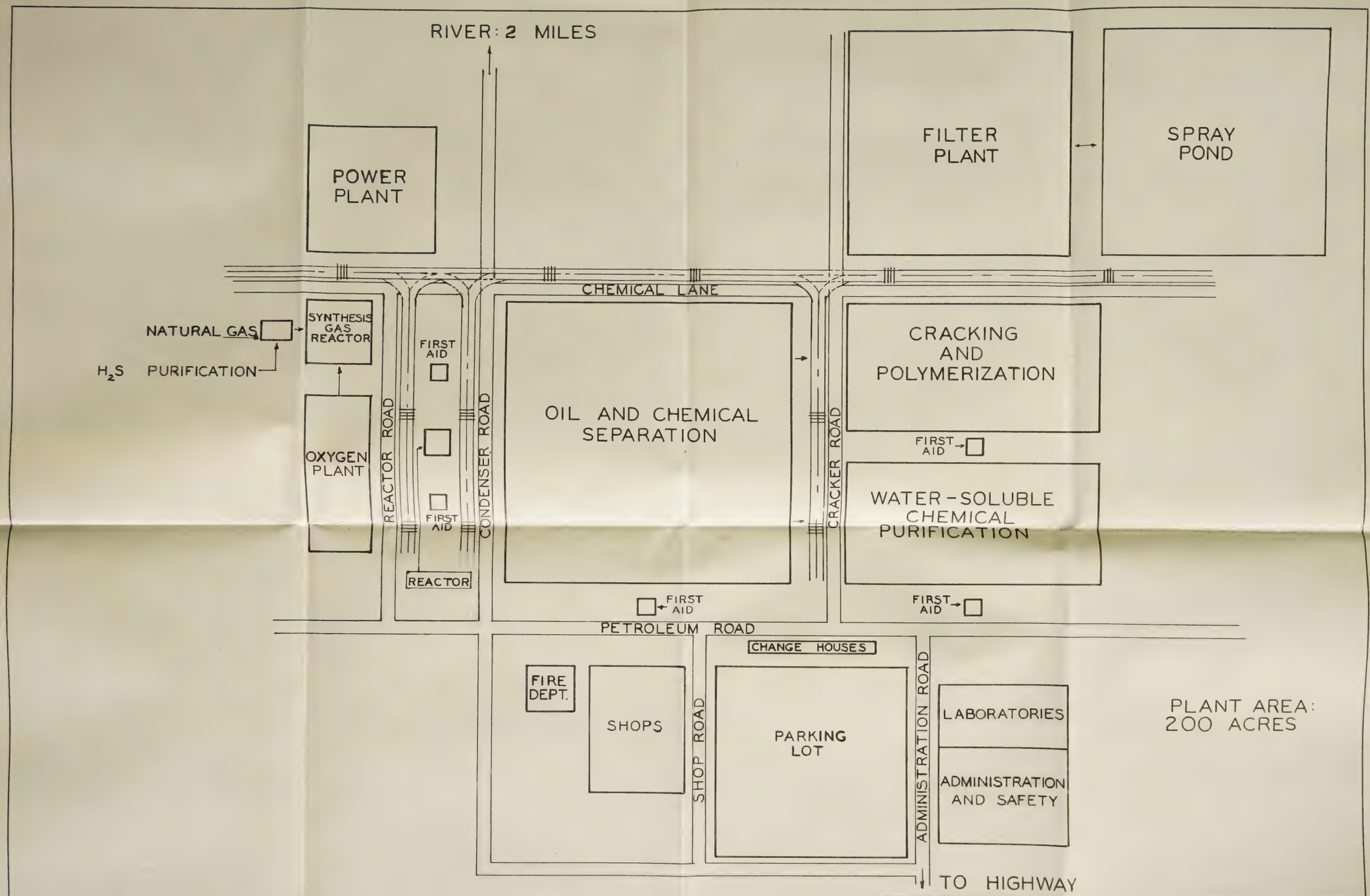
LAYOUT
OF ILLINOIS - URBANA, ILLINOIS
1"=100 5-2-50



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CHECKED BY		[redacted]
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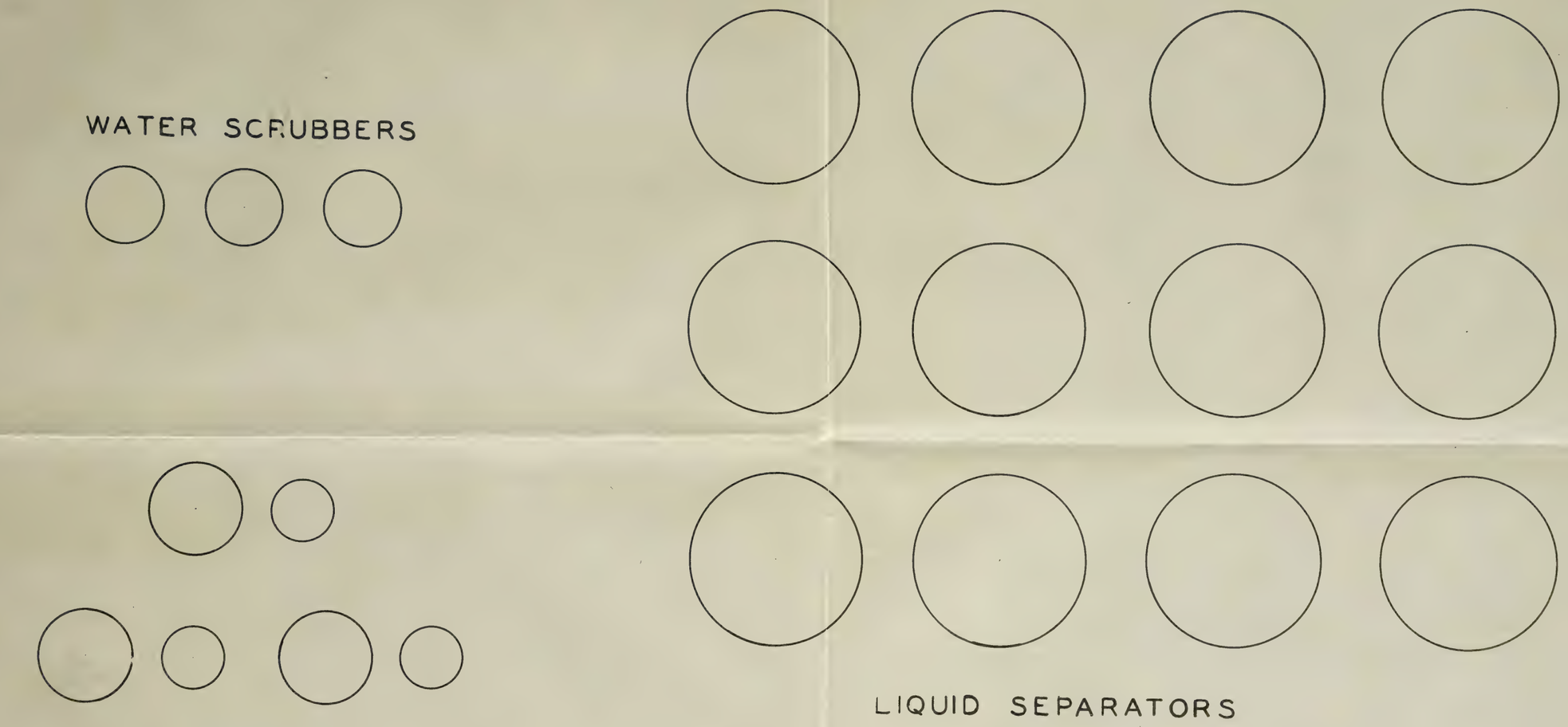
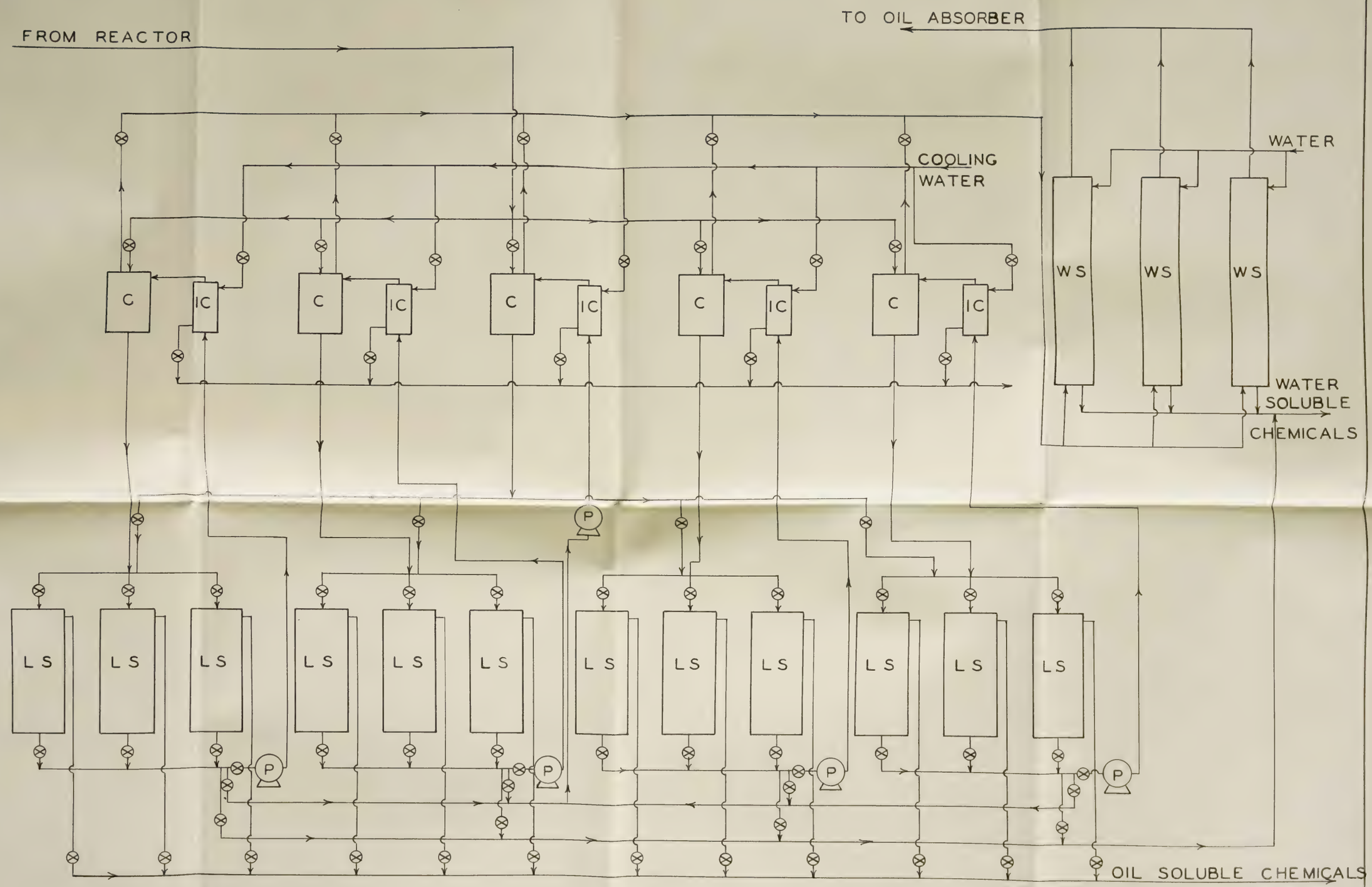
SUBMITTED		SCALE 1" = 100' / DATE 5-21-50
EXAMINED		PART NUMBER
APPROVED		

WATER SCRUBBERS



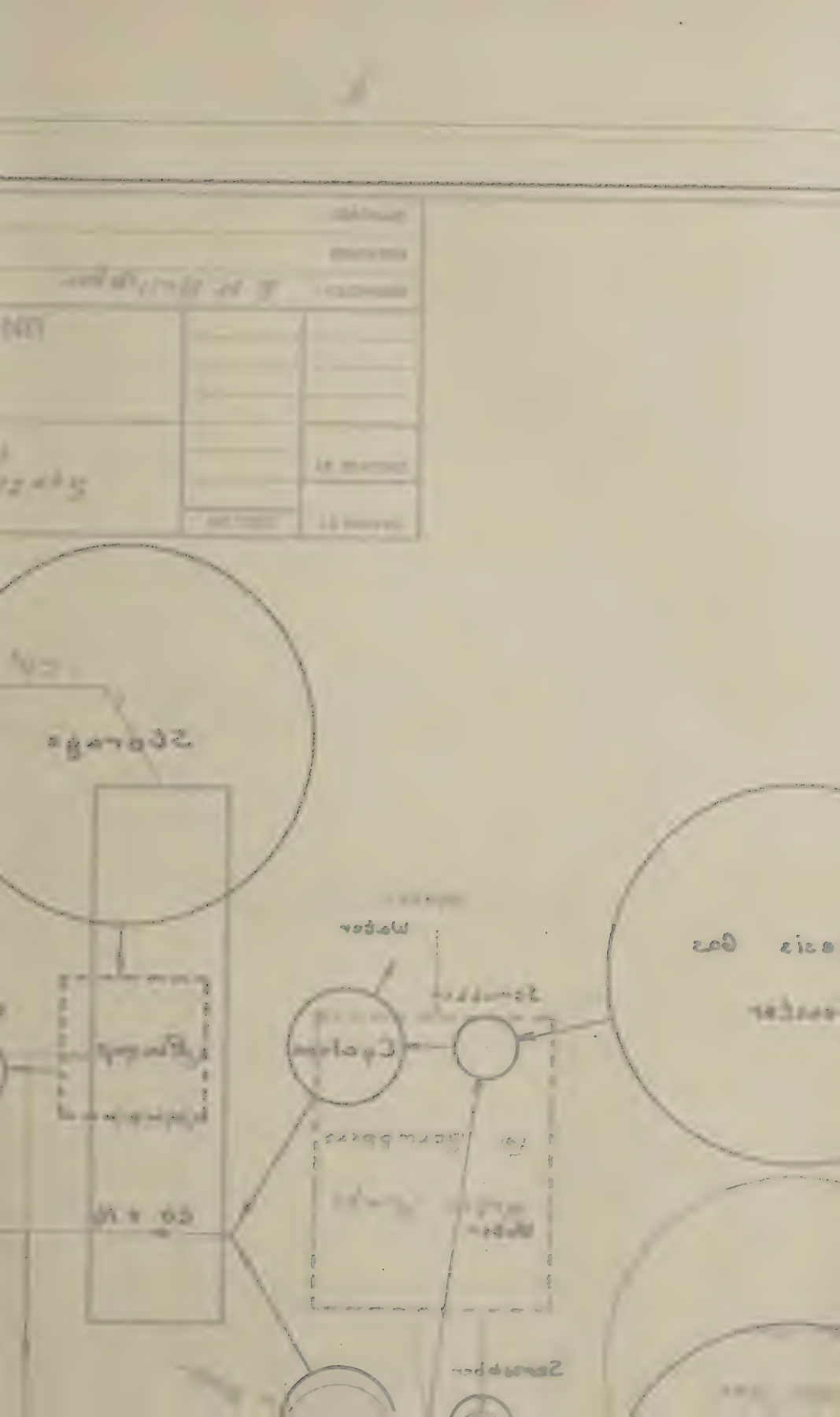
DRAWN BY	USED ON	PRE-PILOT PLANT DESIGN CHEMICAL SEPARATION LAYOUT
CHECKED BY		[redacted] UNIVERSITY OF ILLINOIS - URBANA, ILLINOIS
SUBMITTED		SCALE 1" = 17' / DATE
EXAMINED		
APPROVED		PART NUMBER

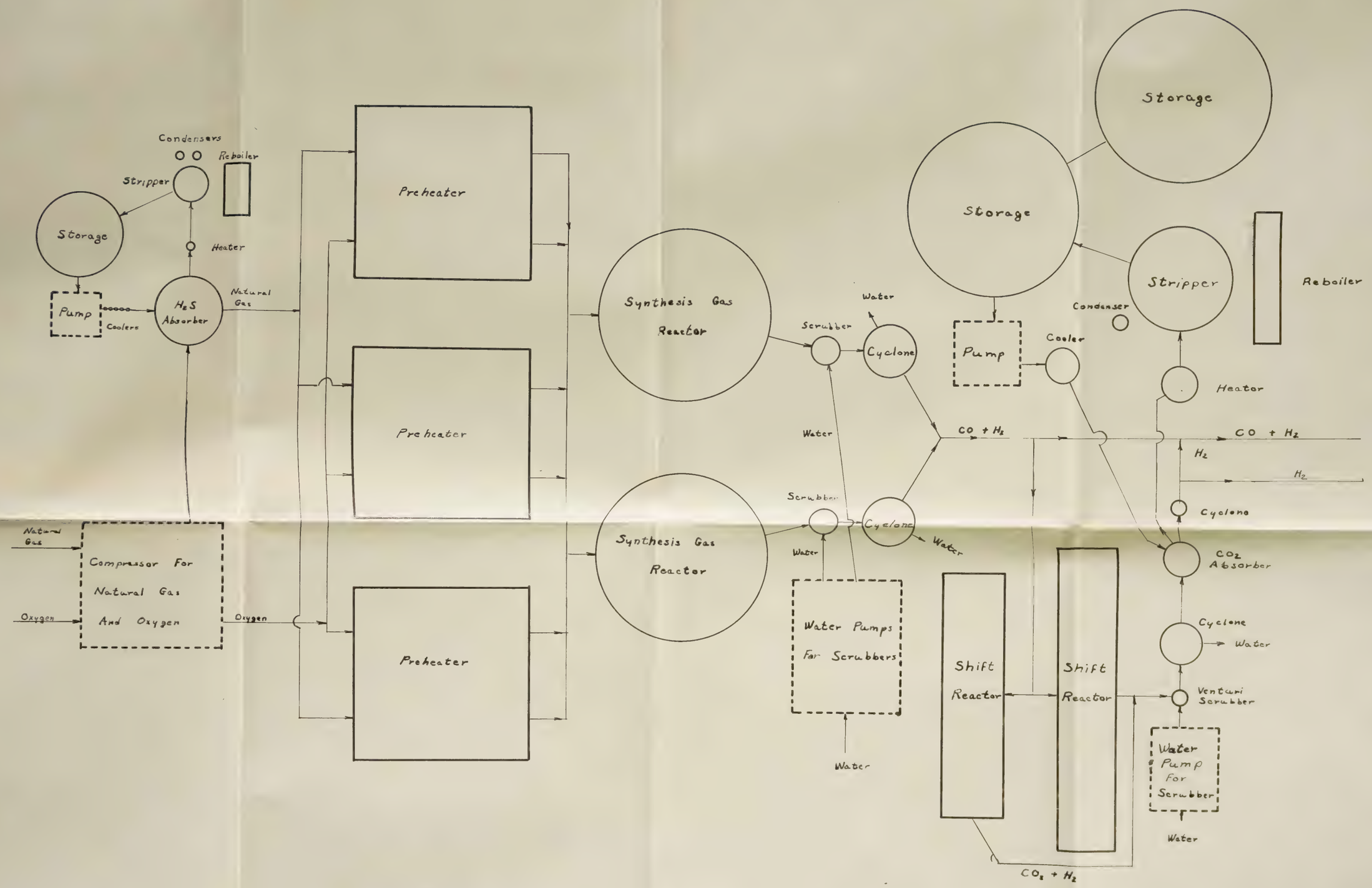
TO OIL ABSORBE
COOLING
WATER
IC
C
IC
C



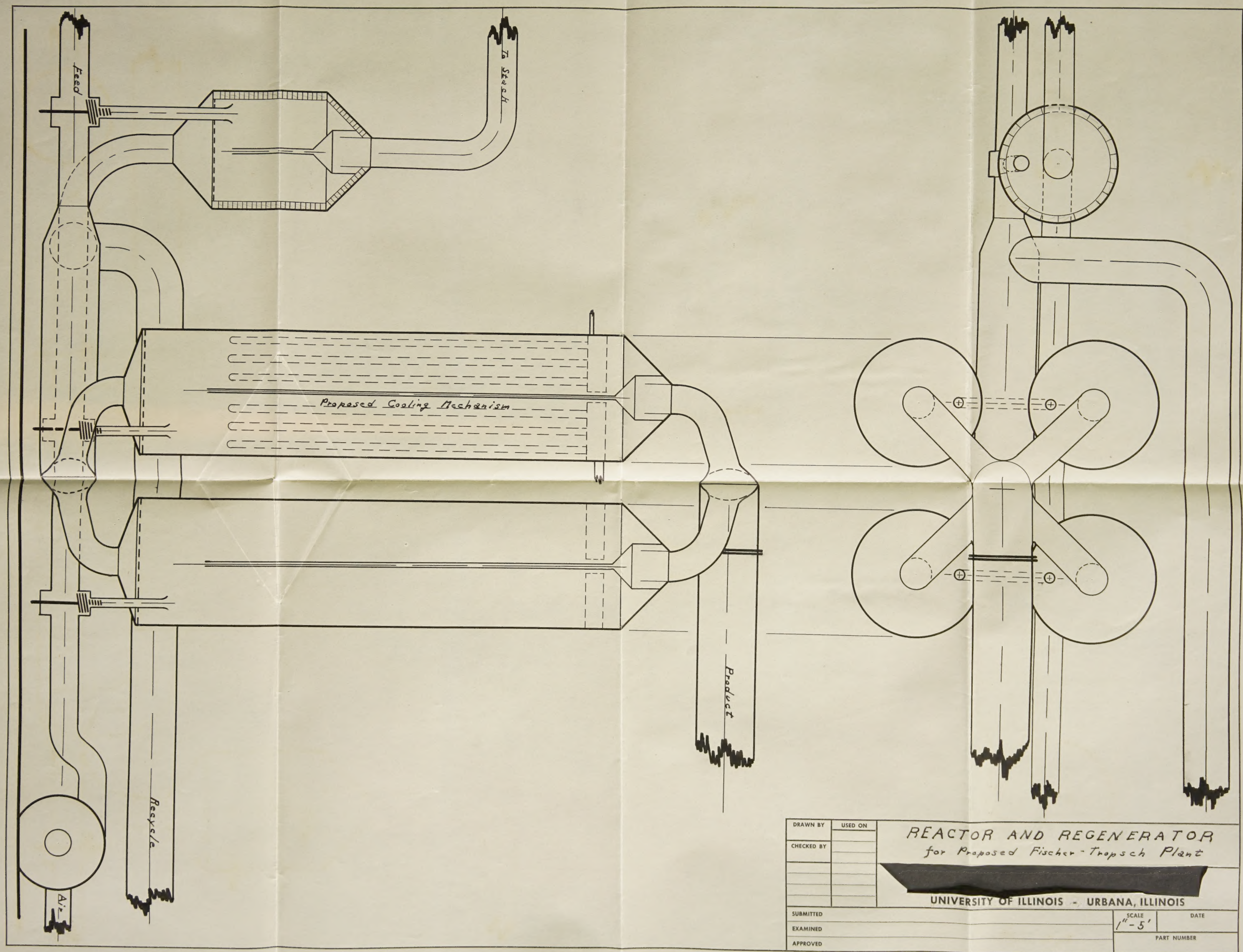
KEY: C—CONDENSER
P—PUMP
IC—INTERCOOLER
WS—WATER SCRUBBER
LS—LIQUID SEPARATOR

DRAWN BY	USED ON	PRE-PILOT PLANT DESIGN CHEMICAL SEPARATION FLOW SHEET
CHECKED BY		[redacted]
		UNIVERSITY OF ILLINOIS - URBANA, ILLINOIS
SUBMITTED		SCALE / DATE
EXAMINED		
APPROVED		PART NUMBER





DRAWN BY	USED ON	Synthesis Gas Section Fischer-Tropsch Pre Pilot Plant Design
CHECKED BY		[redacted]
		UNIVERSITY OF ILLINOIS - URBANA, ILLINOIS
SUBMITTED	E. H. Bollinger	SCALE 1" = 8' / DATE
EXAMINED		PART NUMBER
APPROVED		



DRAWN BY	USED ON	REACTOR AND REGENERATOR for Proposed Fischer-Tropsch Plant
CHECKED BY		[redacted]
		UNIVERSITY OF ILLINOIS - URBANA, ILLINOIS
SUBMITTED		SCALE 1"-5' / DATE
EXAMINED		PART NUMBER
APPROVED		

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